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Modeling of SO2 Removal in Spray-Dryer Flue-Gas Desulfurization System

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MODELING OF SO_2 REMOVAL IN SPRAY-DRYER FLUE-GAS DESULFURIZATION SYSTEM

by

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This report presents a comprehensive mathematical model of the SO_2 removal process in a spray-dryer flue-gas desulfurization system. Simultaneous evaporation of a sorbent droplet and absorption/reaction of SO_2 in the droplet are described by the corresponding heat- and mass-transfer rate relations. Dissolution kinetics of lime particles within a slurry droplet is included in determining the overall SO_2 removal rate. The model identifies several important parameters which need to be estimated or determined from experimental data. This report also includes a computer program, "SPRAYMOD," written in Basic language, to predict SO_2 removal in a spray dryer, based on the model developed. The program is user oriented and easy to use. The contribution of particulate collection equipment, a baghouse and an electrostatic precipitator, toward overall SO_2 removal is also discussed.

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SECTION 1

INTRODUCTION

Spray drying technology for SO_2 absorption/removal from flue gases has been advanced for the past few years. In spite of a large amount of pilot-plant testing and a few full-scale commercial applications, however, there is still a lack of comprehensive predictive modeling of this process. A review of qualitative mechanisms so far proposed was published recently (Apple and Kelly, 1982; Getler et al., 1979). Semiempirical relationships have been developed to relate SO_2 removal efficiency of the spray dryer system with stoichiometric ratio and approach to saturation (Downs et al., 1980; Samuel et al., 1983). Such relationships, due to the empirical parameters, tend to be specific for the spray dryer system used to obtain them. They also consider a few variables and, thus, have a limited use to optimize or troubleshoot a given system.

This report presents a simple mathematical model describing various processes occurring in a spray dryer flue gas desulfurization (FGD) system. The overall process is subdivided into subprocesses contributing to the SO_2 removal. Since this report represents an initial effort of comprehensive modeling of the spray dryer FGD system, attention was given to the simplicity of modeling. Various parameters required for such a model are identified. An overall qualitative picture is first presented, followed by modeling of the different subprocesses. In actual practice, the particulate collection equipment associated with the spray dryer has also been observed to contribute to some extent to overall SO_2 removal. Therefore, the contribution of particulate collection equipment to overall SO_2 removal is also considered in this modeling endeavor.

SECTION 2

CONCLUSIONS AND RECOMMENDATIONS

1. A comprehensive mathematical model was developed to describe various subprocesses occurring in the SO_2 removal process in spray dryer flue-gas desulfurization system. SO_2 removal, both in the spray chamber and in the particulate collection device, was considered.

2. A computer program, "SPRAYMOD," was written to predict SO_2 removal in a spray dryer based on the mathematical model developed in this report.

3. Key physical parameters influencing the model were identified. Several of these parameters, such as critical and equilibrium moisture contents and the reaction coefficient, need to be determined experimentally. These physical parameters, along with the process operating parameters such as stoichiometric ratio and approach to saturation, etc., determine the spray dryer FGD performance.

4. Comparison of model predictions with one set of pilot-plant data shows good agreement. The main obstacle in such comparisons is lack of availability of complete data and the fundamental parameters needed in the model. Future work, therefore, should include procurement of complete data and determination of unknown parameters to allow further validation of the model.

SECTION 3

OVERALL PROCESS

In spray dryer SO_2 removal systems, a conventional spray dryer is typically used to contact SO_2-laden flue gas with spray droplets of a slurry or a solution of a suitable sorbent. A typical schematic of a spray dryer system is shown in Figure 1. Rotary or pneumatic atomizers are used to inject the sorbent slurry/solution. Amount of sorbent added depends upon stoichiometric ratio to be used and the inlet flue gas SO_2 concentration. Amount of water added into the system is controlled by inlet flue gas temperature and humidity and the desired approach to saturation at the spray dryer outlet. Lime slurry is typically prepared in a slaker to obtain a slurry of fine-grained lime particles. Flue gas is usually introduced at the top of spray dryer cocurrently with the sorbent, but other configurations are also possible. The flue-gas residence time is typically about 10 seconds. After the spray dryer, the flue gases, along with flyash and dried sorbent/product particles, pass through particulate control equipment, such as a baghouse or an electrostatic precipitator. Some particulate collection may also occur in the spray dryer itself.

In the spray chamber, two processes occur simultaneously: water evaporates from the droplet and SO_2 is absorbed in and reacts with the alkaline sorbent. The flue gas is typically humidified adiabatically to within 20° to 60°F of its saturation temperature. The sorbent in the spray droplets, along with the reaction products, is evaporated to apparent dryness up to a certain equilibrium moisture content and entrained in the flue gas. The resulting solid particles are removed in a particulate collection device. Typical mean residence times for the slurry droplets in the spray dryer are 8-12 seconds. The amount of water evaporated from a droplet is set by the operating conditions of the dryer-inlet gas temperature, inlet gas relative humidity, system pressure, approach to saturation temperature, and corresponding equilibrium moisture content of the solid. SO_2 is removed by the sorbent both during and after drying of droplets. It is commonly believed that most of the SO_2 is removed before the slurry

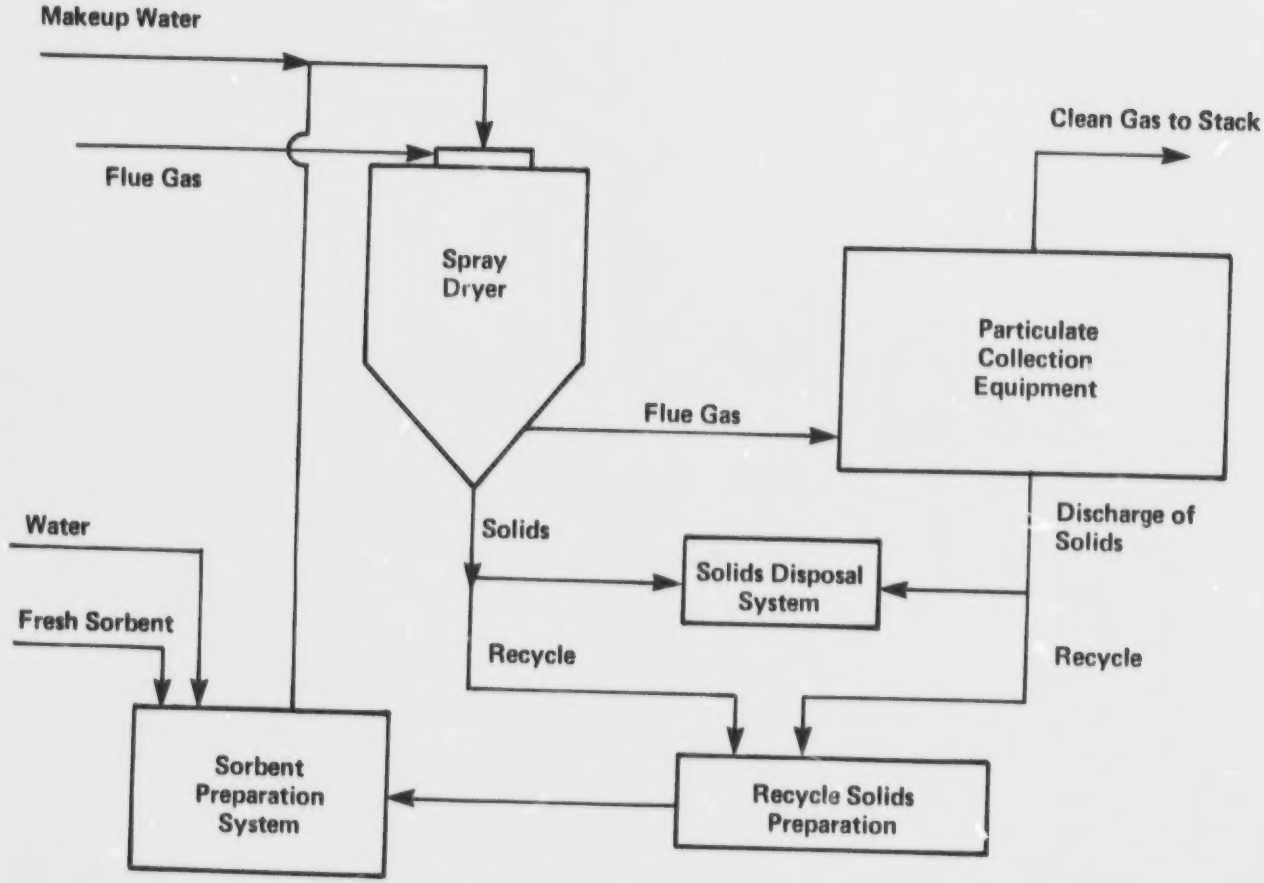


Figure 1. Simplified schematic of a spray dryer FGD system.

droplets are dried. Spray dryer operation nearer to flue-gas saturation condition and higher stoichiometric ratios improve SO_2 removal efficiencies. SO_2 removal may continue through the particulate collection equipment as gas passes through the filter cake in the baghouse or over the deposits on the collection plates in case of an electrostatic precipitator. Further, a portion of the waste particulate discharge from the spray dryer and the particulate collection device may be recycled into the spray dryer's feed slurry to increase the sorbent utilization.

Although droplet evaporation and SO_2 absorption occur simultaneously, the droplet drying process is more or less independent of the SO_2 absorption process. On the other hand, the SO_2 absorption/reaction process has been observed to be strongly related to the drying process and droplet moisture content. For example, closer approach to flue-gas saturation increases the drying time of a droplet and improves SO_2 removal efficiency (Samuel et al., 1983; Stevens et al., 1984). Therefore, in this report the drying of a droplet is considered first to estimate the rate of drying and the droplet moisture content with time, and the total drying time of a droplet. The absorption and reaction of SO_2 in the droplet is then analyzed to determine the rate of absorption/reaction and the SO_2 removal during the droplet drying process. SO_2 absorption/reaction occurs both before and after drying of the droplets in the spray chamber and continues through the particulate collection device. SO_2 removal during different drying stages within the spray chamber and that in the particulate collection equipment are considered separately and treated as subprocesses.

Hydrated lime is predominantly used as a sorbent in spray absorption systems. Some studies have also been conducted with sodium carbonate and bicarbonate which are highly soluble. Since lime, as well as calcium products, are sparingly soluble in water, lime is injected in the form of slurry and a solid phase is present throughout the residence time of the droplet. On the other hand, since the sodium sorbents and products are highly soluble, the sodium sorbents are introduced as solution, and the solid phase does not appear in the droplet until late stages of drying. SO_2 absorption in a droplet of both slurry and solution is considered in this report.

SECTION 4

DRYING OF DROPLETS

When sorbent is introduced in the slurry form, the slurry feed to the spray absorber is usually 60-95 percent moisture and contains finely divided, solid, sorbent particles suspended in a liquid medium. The drying behavior of a slurry droplet with freely moving sorbent particles may be expected to be similar to that of a solution droplet. In line with conventional drying theory (Perry and Chilton, 1973), the evaporation from a slurry/solution droplet may be expected to proceed in two stages. In the first stage, the solid's concentration does not affect evaporation rate, and this stage continues until the moisture level falls below a critical moisture content. During this stage, the rate of evaporation is solely determined by the resistance of the gas film surrounding the droplet to the transfer of water vapor. This is similar to the conventional constant drying-rate condition. In case of a solution droplet, the solid phase appears during this stage due to solubility limitations on the reduced amount of water remaining in the droplet. In the second stage, the solid's concentration reduces the rate of drying, since the moisture now has to diffuse through the solid matrix. During this stage, there is a change from water as a continuous phase, as initially in the droplet, to the solid matrix as a continuous phase; the solid individual particles touch each other and are no longer mobile in the droplet. The drying continues until the droplet moisture content approaches an equilibrium moisture content in equilibrium with the surrounding gas atmosphere. It is important to determine how the solid's concentration affects the drying rate in this period as it determines the overall drying time of a droplet. The drying behavior of droplets in a spray chamber has been studied extensively, at least for the first stage (Sjenitzer, 1952; Kerkhof and Schoeber, 1974; Marshall, 1955; Gauvin and Katta, 1976; Crowe, 1980). The drying processes in the two stages are considered in the following section.

4.1 CONSTANT RATE OF DRYING PERIOD

The drying rate in the first stage is determined by the simultaneous heat transfer from gas phase to the droplet, and mass transfer from the droplet to the gas phase. The transfer coefficients may be determined by a rigorous analysis for a quiescent droplet-gas system or from correlations such as (Ranz and Marshall, 1952):

$$Sh = 2 + \alpha \ Re^{\beta} Sc^{\gamma} ,$$

$$Nu = 2 + \alpha \ Re^{\beta} Pr^{\gamma} , \qquad (1)$$

where Sh is Sherwood Number for mass transfer ($k_d d_d/D$); Re is Reynolds Number based on droplet diameter; Sc is Schmidt Number ($\mu/\rho D$); k_d is gas-phase mass transfer coefficient; d_d is droplet diameter assuming a spherical droplet; D is gas-phase diffusivity; μ is gas-phase viscosity; ρ is gas-phase density; Nu is Nusselt Number for heat transfer (hd_d/k); Pr is Prandtl Number ($C_p \mu/k$); h is gas-phase heat transfer coefficient; k is gas-phase thermal conductivity; and C_p is gas-phase specific heat. The best fit values for constants α, β, and γ, as given by Ranz and Marshall (1952), appear to be 0.6, 0.5, and 0.33, respectively. The Reynolds Number represents the influence of the relative velocity between gas phase and droplet, and based on the droplet diameter (assuming a spherical droplet) is given by:

$$Re = \frac{d_d v \rho}{\mu} , \qquad (2)$$

where, v is the relative velocity between gas phase and droplet.

The movement of droplets in a spray chamber is highly complicated, but may be broken down into two stages: 1) deceleration, and 2) free fall, with respect to the gas phase. Droplets often have very high initial velocity leaving the atomizer and in the first stage are decelerated due to the resistance of the surrounding gas to a steady-state free-fall velocity. In the second stage, the droplet is considered to continue to move at the free-fall velocity. The deceleration time depends upon the droplet diameter and is usually very short. For example, a 100-μm droplet may be decelerated

from 67 m/sec to 5 m/sec within 0.0185 seconds (Sjenitzer, 1952). For smaller droplets even lower time would be required for the same deceleration. To take into account the deceleration stage and variation in droplet Reynolds Number would require solving the momentum balance equation for the droplet. Since this period is usually of short duration (less than 0.1 second), its contribution to drying of the droplet and SO_2 removal may be ignored in the present simple modeling.

The free-fall velocity of the droplets in the gas phase may be given by Stokes' Law as:

$$v = \frac{(\rho_1 - \rho)\ g\ d_d^2}{18\ \mu}\ , \qquad (3)$$

where, ρ_1 is the density of droplet; and g is acceleration due to gravity.

For a 50-μm water droplet, the free-fall velocity is about 6.8 cm/sec and the corresponding Re is about 0.2. The Sherwood Number, as given by Equation (1), is about 2.2. Thus, the contribution of the relative velocity to the Sherwood Number [or Nusselt Number (Nu)] for a 50-μm droplet is about 10%, and for smaller droplets the effect of relative velocity will be less. The relative free-fall velocity, as given by Equation (3), is obtained under laminar flow conditions. The gas-phase turbulence and swirling motion, imparted due to tangential gas velocity, tends to entrain the droplets and reduce the relative velocity. It is extremely difficult to estimate the effect of gas-phase turbulence. For the present modeling, Equation (3) may be used to obtain relative velocity. Equation (1) is then used to predict the dimensionless numbers Sh and Nu. If the effect of relative velocity is ignored, Equation (1) simplifies to:

$$Sh,\ Nu \approx 2\ . \qquad (4)$$

The gas-phase mass transfer coefficient, k_d, is then given by:

$$k_d = \frac{2\ D_{water\text{-}gas}}{d_d}\ , \qquad (5)$$

and the heat transfer coefficient, h, is given by:

$$h = \frac{2\,k}{d_d} \quad . \tag{6}$$

For a droplet temperature, T_d; bulk gas temperature, T_g; bulk water vapor mole fraction, $x_{g,w}$; and equilibrium vapor-mole fraction at droplet surface, $x_{d,w}$ (see Figure 2a); the respective rates of mass and heat transfer are given by (assuming negligible convective transport):

$$N_w = k_d\, \pi d_d^2\, c\, (x_{d,w} - x_{g,w}) \quad , \tag{7}$$

and

$$Q = h\, \pi d_d^2\, (T_g - T_d) \quad , \tag{8}$$

where c is mean molar density of the gas phase; N_w is the molar rate of water evaporation; and Q is [illegible] of heat transfer from gas-phase to droplet.

The heat transferred from gas phase to the droplet is used to evaporate the moisture; thus, the heat and mass transfer are related by:

$$Q = N_w \lambda + \frac{\pi d_d^3}{6}\, \rho_1\, C_{p,1}\, \frac{dT_d}{dt} \quad , \tag{9}$$

where ρ_1 and $C_{p,1}$ are density and specific heat of the droplet, and λ is the molar heat of vaporization for water.

Also, equilibrium conditions prevail at the droplet surface; therefore:

$$x_{d,w} = f(T_d) \quad , \tag{10}$$

as given by the water vapor pressure at T_d.

Similar correlations are also obtained following a rigorous analysis of the evaporation process, assuming quiescent droplet-gas system (Figure 2b). The resulting relations for rate of evaporation and heat transfer are given as (Bird et al., 1960):

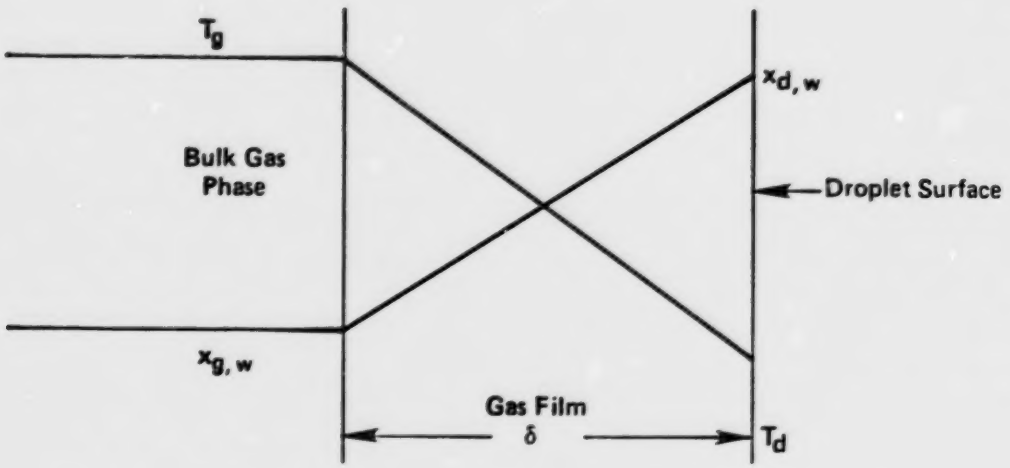


Figure 2a. Gas phase concentration and temperature gradients near a droplet.

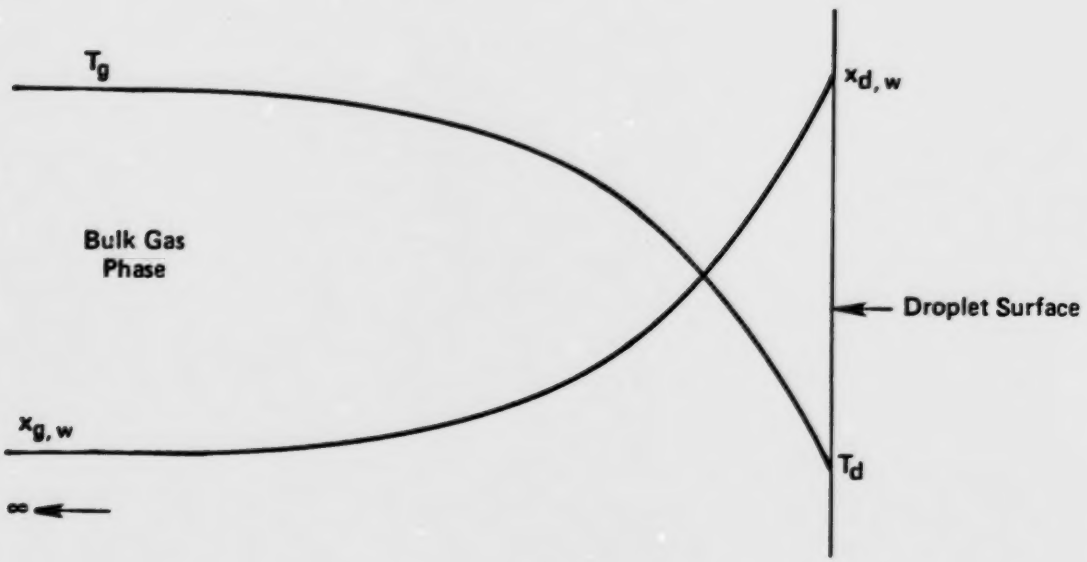


Figure 2b. Schematic of quiescent droplet-gas system.

$$N_w = 2\pi\, D_{water\text{-}air}\, c\, d_d \ln\left[\frac{1 - x_{g,w}}{1 - x_{d,w}}\right], \qquad (11)$$

and

$$Q = 2\pi\, k\, d_d\, (T_g - T_d)\, \left(\frac{\varepsilon}{e^{\varepsilon} - 1}\right), \qquad (12)$$

where ε is dimensionless heat-transfer rate factor, and for water is given by:

$$\varepsilon = \frac{9}{\pi} \frac{N_w C_{p,1}}{k\, d_d}, \qquad (13)$$

where $C_{p,1}$ is specific heat of water vapor.

Again, the heat balance condition and droplet surface equilibrium condition given by Equations (9) and (10) apply.

Equations (11) and (12) take into account the effect of evaporation of water on the rates of heat and mass transfer, and they approach Equations (7) and (8) for low droplet Reynolds Number when Sh, Nu $\approx$ 2, and for low rates of water evaporation. Equation (12) indicates that evaporation of water reduces heat flux from the gas phase to the droplet.

4.2 FALLING RATE OF DRYING PERIOD

The constant rate-drying process continues until the moisture content of the droplet falls below a critical moisture content at which the solid's concentration begins to influence the drying rate. The exact nature of this influence, as given by a plot of rate of drying vs. moisture content, depends upon the solids content of the slurry (the reagent and reaction products). Critical moisture content of the droplet is also a property peculiar to its solids properties such as hygroscopicity (van Brakel, 1980). The critical moisture content may be considered as that at which the solid particles begin to touch each other and form a continuous phase. The drop diameter then does not change during further drying (Parti and Palancz, 1974). For spherical particles arranged in a cubical pattern, the

corresponding critical moisture content at which the solid spheres touch is 47.6%.

Usually, the falling rate is characterized by two modes (Masters, 1976). In the first mode, the drying proceeds by unsaturated surface drying due to the increasing proportion of dry spots on the surface. This mode is characterized by a falling rate which is linearly proportional to the moisture content. Since a slurry droplet is continuously in motion and the primary particles within a droplet are free to move in the constant rate period, the entire surface of the droplet may be expected to dry at the same time. Thus, the unsaturated surface-drying mode may be expected to be missing completely in the present case.

In the second mode, drying is controlled by diffusion of moisture through the solid matrix. No simple correlations are available to describe the drying rate in this mode, but may be assumed to fall linearly between the critical-moisture content and equilibrium-moisture content. The second mode continues till the moisture content reaches an equilibrium-moisture level. The drying rate in this mode may be considered to be of the form:

$$\text{Rate of Drying} = \begin{pmatrix} \text{Rate of drying based on} \\ \text{gas-phase resistance alone} \end{pmatrix} \left(\frac{X - X_e}{X_c - X_e} \right)^n , \quad (14)$$

where X is moisture content at time t, weight percent dry basis; X_e is equilibrium-moisture content, weight percent dry basis; X_c is critical-moisture content, weight percent dry basis; and n is a dimensionless power and is equal to 1 for a simple linear relationship. The parameters, critical-moisture content and equilibrium-moisture content, depend upon the droplet's solids content and their properties and must be determined experimentally. The critical-moisture content for lime, clay, etc., is typically in the range of 25-60 percent moisture. Solids properties, such as hygroscopicity and deliquescent nature, and the relative humidity of surrounding gas determine the equilibrium-moisture content. For given solids, equilibrium-moisture content is linear with respect to relative humidity of the surrounding gas phase. In case of highly soluble sorbents and products, their equilibrium solubility should be taken into account in determining equilibrium-moisture content. For highly soluble and hygroscopic

sorbents, the falling-rate period may turn out to be very short with the droplet drying mostly in constant-rate period.

Complication may arise during the drying process due to a very high initial rate of evaporation. This may lead to a solid crust formation on the surface with a lot of trapped moisture. This would lead to eventual breakup of the droplet and irregular particles. Such crust formation would also reduce the rate of drying considerably before it actually reaches the critical-moisture content.

In spray-drying operation, most of the droplet drying is complete within 1-2 seconds. This is also essential so as to avoid any sticking of wet droplets/particles on the dryer walls. In this respect, gas-flow pattern and droplet size are also important. As smaller droplets lose their momentum more rapidly, they are likely to stay aligned longer with the gas-flow profiles. Since most of the moisture is lost in the constant-rate drying conditions, this stage should typically last for 1-2 seconds so as to avoid any sticking problems.

There have been no measurements to indicate relative lengths of drying times for the two stages. The final moisture contents in droplets at the spray-dryer exit were reported by Felsvang (1982) from a pilot-plant study and were found to depend upon the approach to saturation temperature and the droplet size. If all the droplets reach an equilibrium with the outlet flue gas, their moisture content would be independent of droplet size and should only depend upon approach to saturation (relative humidity of the gas). The fact that it depends on droplet size indicates that the larger droplets did not reach an equilibrium by the end of spray dryer in this particular study. No droplet size measurements were reported in this study, and only two categories, fine and very fine droplets, were designated. If it is assumed that the "very fine" droplets did reach an equilibrium-moisture content, then a plot of equilibrium-moisture content vs. relative humidity may be constructed as shown in Figure 3. The measured moisture contents in this study ranged from 0.5 to 1.5 percent for "very fine" droplets and 0.5 to 3.0 percent for "fine" droplets. There seem to be many variations in measured values of solids moisture content at the spray-dryer outlet. For example, Stevens et al. (1984) reported measured values of moisture content which ranged from 5-20 percent, which is almost an order of magnitude higher than those in Felsvang's (1982) work. Stevens et al.

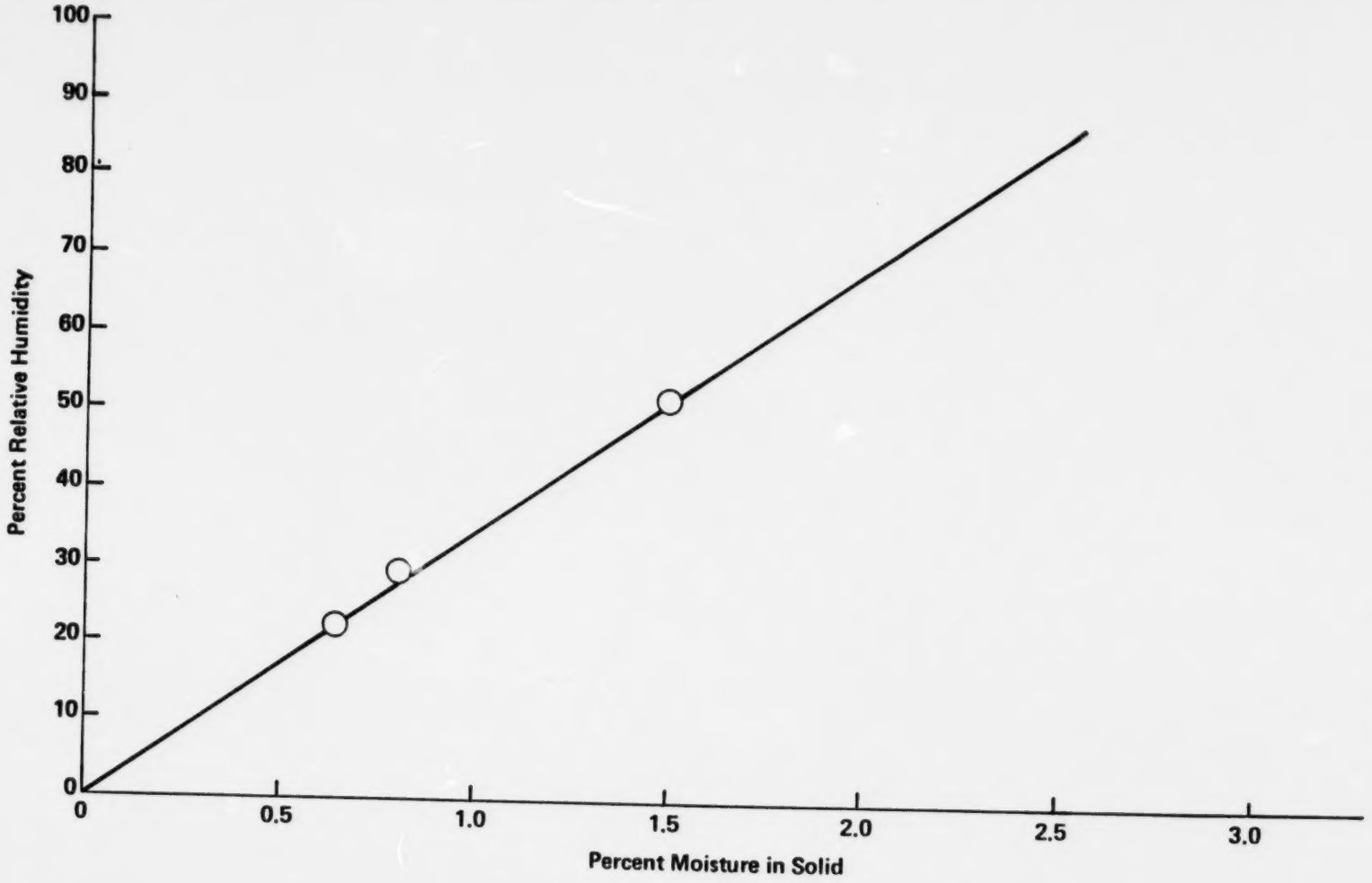


Figure 3. Equilibrium moisture content of solids.

(1984) also reported variations in moisture content with residence time and atomizer speed for the same approach to saturation, indicating that the droplets had not reached an equilibrium-moisture level.

Determination of equilibrium-moisture content is very important to estimate the drying time of the droplet in the second stage. This is also critical for prediction of SO_2 removal, as the aqueous reactions may be assumed to continue until the droplet reaches low enough moisture content. High equilibrium-moisture content may enhance continued SO_2 removal after the droplet is "dried" to the equilibrium.

The critical-moisture content divides the total drying time in the above-mentioned two stages and is thus important as different drying-rate relations apply in each stage.

The drying time required for each stage may thus be determined using the appropriate drying-rate expression and the parameters, critical- and equilibrium-moisture content of solids.

SECTION 5

SO_2 ABSORPTION/REACTION IN SPRAY CHAMBER

Absorption and reaction of SO_2 in a sorbent droplet occurs both before and after drying of the droplet. Presence of moisture during the wet-droplet stage plays an important role as it provides an aqueous medium for absorption and fast ionic reaction of SO_2. The lack of moisture in the dry-particle stage is expected to considerably reduce the rate of SO_2 removal as absorption and reaction in the solid phase is likely to be slower. This is especially true for the lime sorbent which has a low reactivity in solid phase. Sodium reagents are more reactive in solid phase than lime, but the aqueous medium may still improve the rate of SO_2 removal by sodium-based sorbents. The mechanisms of SO_2 removal in the two stages are distinctly different and, therefore, will be considered separately.

5.1 WET-PARTICLE STAGE

During the wet stage, moisture in the droplet participates actively in the overall SO_2 removal process. SO_2 is transported from the bulk-gas phase to the droplet surface in the gas-phase diffusion process. The dissolved SO_2 migrates from the interface to the interior of the droplet by liquid-phase diffusion and reacts with the dissolved sorbent. If a sparingly soluble sorbent, such as lime is used, dissolution of lime will also be important. Lime particles dissolve continuously to match lime depletion in the reaction. If the ionic reaction between sorbent and SO_2 is considered as very fast, both species may migrate to a reaction plane or zone in the bulk liquid as shown in Figure 4. The dissolution process will not be present in case of a soluble sorbent, but the overall process may still be represented by Figure 4 by replacing the equilibrium-solubility concentration by the bulk liquid-phase sorbent concentration.

The above picture is easier to visualize for a wet droplet with a moisture content greater than critical when the sorbent particles are

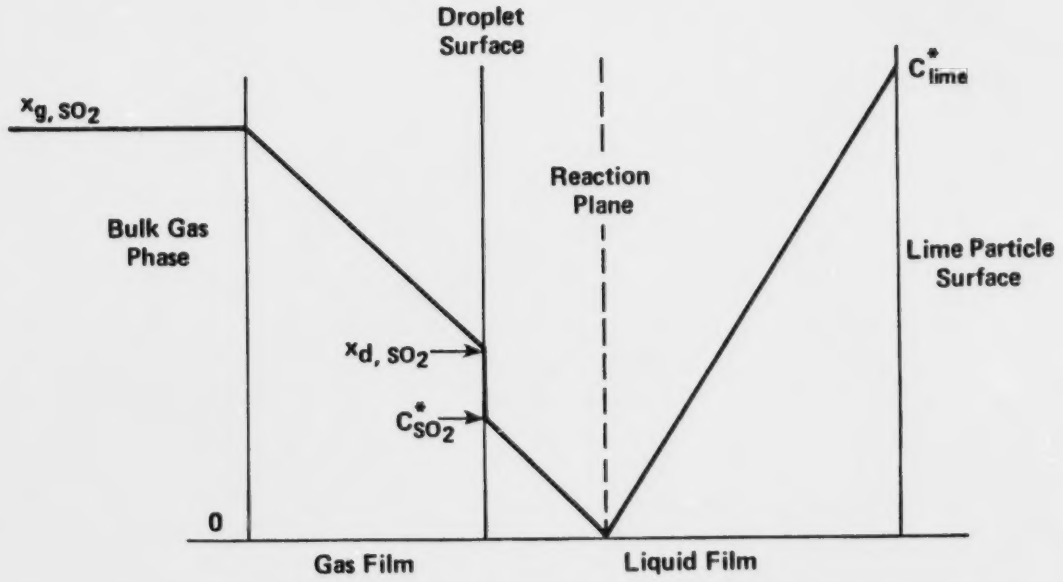


Figure 4. Schematic of SO_2 absorption/reaction in a wet droplet.

freely moving. The above processes become less obvious during further drying as the water phase becomes discontinuous. Nevertheless, for model simplicity, the moisture held within particle interstices may still be assumed to provide enough aqueous medium for the SO_2 absorption process. SO_2 removal during the wet stage by the above mechanism may be assumed to continue till the moisture content in the droplet falls to a certain low level, the lower limit being the equilibrium moisture content. After that, the droplet may be considered as a dry-porous solid to determine further SO_2 removal.

Various basic steps that are involved in the removal of SO_2 during a wet-droplet stage may be described as follows:

1. Transfer of SO_2 from bulk-gas phase to droplet/particle surface. The rate of transfer in this step is controlled by the resistance of a gas film around the droplet.

2. Dissolution of SO_2 into the liquid phase in the droplet, and transfer of dissolved SO_2 from the droplet surface to interior liquid. The transfer rate is controlled by the liquid-film resistance in this step.

3. Dissolution of lime into the liquid phase.

4. Ionic reaction between dissolved SO_2 and dissolved lime in liquid phase.

5. Transfer of reaction products to precipitation sites.

These steps may be expected to have different relative contributions to the overall transport process under different operating conditions and the two distinct drying stages (constant rate and falling rate) described in the preceding section.

In the initial "constant rate" drying stage, water is the continuous phase in the droplet with the primary sorbent particles suspended in it. The particles themselves would be in continuous motion within the droplet. In such a situation, the resistance offered by precipitation of the reaction products may be ignored. Also, the ionic reactions are very fast and may be considered to be not controlling the overall rate of SO_2 removal. The other three transfer processes may be analyzed in detail to determine their relative importance.

5.1.1 Gas-Film Resistance

Diffusion of SO_2 from the bulk-gas phase to the droplet surface is similar to water evaporation from the droplet surface, and similar correlations apply.

The mass transfer Sherwood Number is given by:

$$Sh = 2 + \alpha\, Re^{\beta} Sc^{\gamma} , \qquad (1)$$

and in a limiting case for small droplets and low Reynolds Number (as discussed in previous sections), $Sh \approx 2$. The mass-transfer coefficient, k_d, is then given by:

$$k_d = \frac{2\, D_{SO_2\text{-gas}}}{d_d} . \qquad (5')$$

The rate of SO_2 transfer would be given by:

$$N_s = k_d(\pi d_d^2)\, c\, (x_{g,s} - x_{d,s}) , \qquad (15)$$

where $x_{g,s}$ and $x_{d,s}$ are mole fractions of SO_2 in bulk gas phase and at the droplet surface, respectively.

The above formulation does not however take into account the effect of water evaporation on SO_2 diffusion. This may be done in a rigorous analysis for a quiescent droplet-gas system which gives the rate of SO_2 mass transfer under condition, $N_w >> N_s$, as (Bird et al., 1960):

$$N_s = \frac{x_{g,s} N_w}{(F - 1)} , \qquad (16)$$

for droplet interface SO_2 concentration, $x_{d,s} \approx 0$;

where $F = \exp\,(N_w/2\pi\, D_{SO_2\text{-gas}}\, c\, d_d)$.

For nonzero surface SO_2 concentrations:

$$N_s = \frac{x_{g,s} N_w}{F - 1} - \frac{N_w x_{d,s} F}{F - 1} \quad . \tag{17}$$

The rate of mass transfer, as given by Equations (15)-(17), is strongly dependent on SO_2 concentration in the bulk-gas phase. The derivation of Equations (16) and (17) assume that the molar rate of evaporation of water, N_w, is much greater than molar rate of SO_2 absorption. This is usually true throughout the drying process. If during late drying stages of the falling-rate period N_w becomes comparable with N_s, Equation (15) should be used.

5.1.2 Liquid-Film Resistance

No good correlations are available to estimate the liquid-phase resistance in the droplet to the mass transfer. An order of magnitude estimate may be obtained. For a droplet diameter of 25-50 μm containing primary lime particles of the order of 1-5 μm, the liquid-film thickness, δ, at the surface contributing this resistance may be considered to be of the order of 1 μm. The liquid-phase mass transfer coefficient may then simply be obtained as the ratio of diffusivity of SO_2 in liquid and δ:

$$k_l = \frac{D_{SO_2\text{-water}}}{\delta} \tag{18}$$

$$\approx \frac{1.7 \times 10^{-5} \text{ cm}^2/\text{sec}}{1 \times 10^{-4} \text{ cm}} \approx 0.17 \text{ cm/sec} \quad .$$

Important consideration should be given to the enhancement of this liquid film mass-transfer coefficient due to the very fast reaction of dissolved SO_2 with dissolved lime in the bulk liquid. From the theory of mass transfer with chemical reaction in liquid phase (Danckwerts, 1970), this enhancement may be expected to be on the order of ϕ:

$$\phi = 1 + \frac{D_{\text{lime-water}}\, C^*_{\text{lime}}}{D_{SO_2\text{-water}}\, C^*_{SO_2}} , \qquad (19)$$

where the D's are diffusivities of lime and SO_2 in water, C^*_{lime} is the concentration of lime in the bulk liquid corresponding to its equilibrium solubility, and $C^*_{SO_2}$ is the equilibrium concentration of SO_2 in liquid phase at the droplet surface. (Similar relations also apply for other sorbents.)

The rate of transfer of SO_2 in the liquid phase from surface to bulk liquid is then given as:

$$N_s = k_1\, \phi\, \pi d_d^2\, (C^*_{SO_2} - C_{SO_2,1}) , \qquad (20)$$

where $C^*_{SO_2}$ is the equilibrium SO_2 concentration at the droplet surface, and $C_{SO_2,1}$ is the bulk concentration of SO_2 in the liquid phase.

For fast reaction between dissolved SO_2 and sorbent, both the species diffuse to a reaction front where their concentrations drop to zero (see Figure 4). In this case, $C_{SO_2,1} \approx 0$. Equation (20) then becomes:

$$N_s = k_1\, \phi\, \pi d_d^2\, C^*_{SO_2} . \qquad (20')$$

Equation (20') thus gives the rate of liquid-phase SO_2 transfer.

5.1.3 Dissolution of Lime

This step is obviously required only for sorbents such as lime which are sparingly soluble in water. Rate of dissolution of lime may be approximately estimated using the diffusivity of lime in water, its solubility, and the average interparticle surface distance between lime particles in the droplet. Based on a simple film model (Figure 4), the rate of dissolution, R_p, from a single lime particle of diameter, d_p, is given by:

$$R_p = D_{\text{lime-water}} \pi d_p^2 \left(\frac{C_1^* - C_1}{\delta}\right) , \qquad (21)$$

where D is diffusivity of lime in water; d_p is average primary lime particle diameter; C_1^* is equilibrium solubility concentration of lime in water; C_1 is concentration of lime in bulk solution; and δ is liquid film thickness responsible for lime dissolution and may be approximated to one-half of the interparticle distance.

The primary lime particles may be assumed to be arranged in a cubical pattern in the slurry droplet. For an average particle diameter, d_p, and moisture-volume fraction, w, the film thickness, δ, based on interparticle distance is then given by:

$$\delta = d_p \left(\frac{0.973}{(1-w)^{1/3}} - 1.0 \right) . \qquad (22)$$

For a 50 percent slurry, $\delta \approx 0.22\ d_p$. The above derived film thickness is based on interparticle distance.

For fast reaction between dissolved SO_2 and lime, both the species diffuse to a reaction front where $C_1 \approx 0$. The rate of lime dissolution from all lime particles in a droplet, R_d, assuming a monodisperse lime particle size distribution, is given by:

$$R_d = R_p \times \text{number of particles in a droplet}$$

$$= D_{\text{lime-water}} \frac{\pi d_p^2 C_1^*}{\delta} \frac{d_{do}^3 (1-w_o)}{d_{po}^3} , \qquad (23)$$

where, subscript o represents initial values for droplet diameter, moisture fraction, and lime particle diameter. Derivation of Equation (23) assumes a pure lime particle and ignores any effect of product precipitation on dissolution rate, which is discussed in a later section.

Equations (15), (17), (20'), and (23) thus describe the rate of transfer of reacting species in the three subprocesses.

The gas- and liquid-phase mass transfer coefficients may be combined to obtain an overall mass-transfer resistance based on gas phase.

Rate of diffusion of SO_2 in the gas phase is given by Equation (15):

$$N_s = k_d \, (\pi d_d^2) \, c \, (x_{g,s} - x_{d,s}) \, , \tag{15}$$

and corresponding rate in liquid phase is given by Equation (20'):

$$N_s = k_1 \, \phi \, \pi d_d^2 \, C^*_{SO_2} \, . \tag{20'}$$

The gas- and liquid-phase surface concentrations are related by Henry's Law and solubility constant, H:

$$c \, x_{d,s} = H \, C^*_{SO_2} . \tag{24}$$

Equations (15) and (20') may be combined using Equation (24) to give:

$$N_s = k_o \, \pi d_d^2 \, c \, x_{g,s} \, , \tag{25}$$

where k_o is the overall mass-transfer coefficient based on gas phase and is related to individual mass-transfer coefficients by:

$$\frac{1}{k_o} = \frac{1}{k_d} + \frac{H}{k_1 \phi} \, . \tag{26}$$

The relative values of k_d and $k_1\phi/H$ determine the relative importance of gas-phase to liquid-phase mass-transfer resistances.

For a 50-μm droplet:

$$k_d \approx \frac{2 \times 0.2}{50 \times 10^{-4}} \qquad \text{(from Equation 5', 80°C)}$$

≈ 80 cm/sec ,

$k_l \approx 0.20$ cm/sec , (from Equation 18, 50°C)

$$H \approx 0.054 \frac{\text{gmole/cm}^3 \text{ in gas phase}}{\text{gmole/cm}^3 \text{ in liquid phase}} .$$ (based on solubility of SO_2 in water at 50°C)

Equilibrium Solubility of Lime $\approx 1.35 \times 10^{-5}$ gmoles/cm^3.

Concentration of SO_2 in liquid phase at surface, at 50°C drop temperature, for gas-phase $\approx 7.0 \times 10^{-4} \times 100 \times 10^{-6}$ gmoles/cm^3;

Concentration of 100 ppm at surface $\approx 7.0 \times 10^{-8}$ gmole/cm^3.

Diffusivity of SO_2 and lime in water, D $\approx 2 \times 10^{-5}$ cm^2/sec.

$$\phi \approx 1 + \frac{1.35 \times 10^{-5}}{7.0 \times 10^{-8}}$$ (from Equation 19)

≈ 194

$$\frac{k_l \phi}{H} \approx \frac{0.17 \ 194}{0.054} \approx 719 .$$

Thus,

$$\frac{k_l \phi}{H} >> k_d .$$

The above calculations show that generally the gas-phase resistance is much higher than corresponding liquid-phase resistance, or:

$$k_o \approx k_d . \qquad (27)$$

Thus, the liquid-phase mass transfer is unlikely to become a controlling factor in overall SO_2 removal. The controlling mechanisms of SO_2 removal will be determined by relative values of N_s and R_d as given by Equations (25) and (23) for SO_2 gas-phase mass transfer and lime dissolution rate, respectively. If $N_s > R_d$, the overall SO_2 removal rate will be controlled by lime dissolution rate and given by R_d. On the other hand if $N_s < R_d$, the gas-phase mass-transfer resistance to SO_2 transfer will control its removal rate.

5.1.4 Effect of Product Precipitation

In developing the above relationships, the effect of product formation is completely ignored. The reaction products of lime, e.g., $CaSO_3$, have low solubility and would therefore precipitate out. As this precipitation may take place on existing lime particles, it may affect the lime dissolution rate. In deriving relationships for the rate of lime dissolution, it was assumed that all the particle surface area is available for dissolution. Thus, the reduction in lime dissolution rate would be directly proportional to surface area of a lime particle obstructed by product precipitation. In the early stages of drying, this effect would tend to be minimal because of the mobility of lime particles within a droplet and also due to the low level of product formation. This effect would increase significantly as the solid phase becomes the continuous phase, since the immobility would allow the precipitated product layer to build up on the lime particle surface, thus obstructing its dissolution. A simpler way to account for this effect in the model is to ignore it in the early stages until moisture content falls to the critical moisture content, and in later stages consider the dissolution rate to be proportional to the lime fraction remaining in the solids phase.

5.2 DRY-PARTICLE STAGE

The above-developed rate relations apply until a distinct liquid phase is present in the droplet. As the droplet dries and moisture content approaches the equilibrium moisture level, diffusion of SO_2 into the solid matrix becomes important. Primary sorbent particles form a continuous solid-phase matrix and become immobile. Thus, any remaining active reagent

at the surface will be consumed first by reaction, with the "reaction front" proceeding inward. The resistance to SO_2 mass transfer shifts from the gas and liquid phase to the solid phase. Lime dissolution in whatever remaining moisture is present locally within the droplet would still be important, as this determines the local reactivity in the solid phase. The SO_2 removal process thus now becomes that of SO_2 diffusion through solid matrix with chemical reaction. The moisture content at which a distinct liquid phase ceases to exist is the one at which the vapor pressure of the moisture over the solid becomes less than or equal to that of the surrounding gas-phase water vapor pressure.

Analysis of SO_2 absorption during this dry-particle stage may be approached in two ways:

i) The first approach is to follow a rigorous analysis of diffusion of SO_2 into a spherical solid matrix with chemical reaction. For this analysis, diffusivity of SO_2 in the solid matrix and the reaction coefficient are needed. In the lime-SO_2 reaction, the products have a higher mass as compared to the reagent, and thus the reaction would tend to expand the particle. At the beginning of this dry-particle stage, the dried particle may be expected to be very porous due to the fine size of primary lime particles. The expansion of the solid mass due to reaction would thus tend to reduce the porosity and thus would reduce the diffusivity of SO_2 as well. Therefore, the diffusivity of the SO_2 in the solid matrix would be dependent on the local extent of reaction. Also, unknown is the reaction coefficient. From a number of pilot-plant experiments, this coefficient is found to be strongly dependent on the solids moisture content (Wilkinson and Tonn, 1982). Moisture content helps lime dissolution and promotes ionic reaction. The graininess of the dried particle or the size of the primary lime particles is also expected to be important in determining the reaction coefficient. Smaller lime particles provide a larger surface area for reaction and would, thus, enhance the reaction coefficient. The local reaction rate would basically depend upon the local SO_2 concentration, reaction coefficient, and local mass fraction of lime in the solid.

Mathematically, the SO_2 diffusion into a solid sphere with chemical reaction may be described by (Crank, 1964):

$$\frac{\partial C}{\partial t} = D_{eff,r} \frac{\partial^2 C}{\partial r^2} + \frac{2}{r} \frac{\partial C}{\partial r} - K_r C C_{lime,r} , \qquad (28)$$

where C is gas-phase concentration of SO_2 in the solid matrix at a radial position, r, at time, t; $D_{eff,r}$ is effective diffusivity of SO_2 in solid matrix. (Since it depends on local extent of reaction, it would also be a function of radial position, r, within a particle); K_r is volumetric reaction coefficient; and $C_{lime,r}$ is concentration of active lime (sorbent) in the solid matrix. The molar rate of SO_2 removal by the dried particle, N_s, is given by the amount of SO_2 consumed by reaction throughout the solid matrix and is also equal to the SO_2 diffusion rate at the surface:

$$N_s = \int_0^{d_d/2} K_r C C_{lime,r} 4\pi r^2 \, dr , \qquad (29)$$

$$= D_{eff,R} \left. \frac{\partial C}{\partial r} \right|_R \pi d_d^2 . \qquad (30)$$

The boundary and initial conditions necessary to solve Equation (28) are:

$$\frac{\partial C}{\partial r} = 0 \qquad \text{at } r = 0 \text{ for all } t;$$

$$C = C_{SO_2,g} \qquad r = d_d/2 \text{ for all } t;$$

$$\text{at } t = 0 \qquad C = C_{SO_2,g} \text{ at } r = d_d/2;$$

$$C = 0 \text{ for } r < d_d/2.$$

The above equations may be solved by numerical methods to determine the amount of SO_2 removed in a given time. Parameters needed for this solution are the effective diffusivity and the reaction coefficient. The gas-phase

SO_2 concentration, $C_{SO_2,g}$, depends upon the rate of SO_2 removal by the dried particles and is determined by overall gas-phase mass balance.

The above equation may be simplified for cases in which either the diffusion process in the solid matrix or the chemical reaction in the solid matrix is dominant over the other. When the chemical reaction rate is relatively much greater than the diffusion coefficient, the SO_2 is consumed by the solid phase as soon as the SO_2 diffuses inward. This leads to a sharp reaction front proceeding inward, and the SO_2 concentration profile in the solid phase appears as a square wave moving inward. The rate of advancement of this reaction front is determined by the chemical reaction rate and is easy to derive. At the other extreme, when the diffusion in the solid matrix is relatively much faster than the rate of chemical reaction, the SO_2 concentration throughout the particle will be uniform and equal to the gas-phase concentration. The reaction will then proceed throughout the volume of the particle. This approach is considered below.

ii) A simple approach is to use a bulk-volume reactivity (K_r) for the entire droplet/particle. This reaction coefficient would depend upon the moisture content of the droplet/particle, primary lime particle size or available surface area, and the diffusivity of SO_2 in the solid material. In a given dry-scrubbing application, the solid reagent, its source, and preparation method are pretty much set, and this leaves only the moisture content of the droplet/particle as a variable process parameter. Thus, for a given system, the overall volume reactivity of a droplet/particle may be approximately expected to be dependent on moisture content alone.

The rate of SO_2 removal by a single droplet/particle with this approach is simply given by integration of Equation (29):

$$N_s = K_r \left(\frac{\pi d_d^3}{6}\right) C_{SO_2,g} \, C_{\text{remaining lime}} \,, \qquad (31)$$

where K_r = f (moisture content). Such a correlation between K_r and moisture content may easily be obtained by well-designed experiments. This approach is thus very simple in mathematics and may also prove practical to use. The gas-phase SO_2 concentration, $C_{SO_2,g}$, depends upon the rate of SO_2

removal by the dried particles and is determined by the overall gas-phase mass balance. The concentration of remaining lime with time is determined by the above rate relation and mass balance over the droplet.

SECTION 6

SPRAY-DRYER INLET AND OPERATING PARAMETERS

The rate relations developed in Sections 4 and 5 describe the mass- and heat-transfer interactions between a single droplet/particle and the surrounding gas phase. The rate of drying and SO_2 removal are interrelated to bulk-gas and droplet properties. In addition, the bulk-gas properties also depend upon gas-inlet conditions, prescribed operating conditions, and gas flow and mixing within a spray chamber.

6.1 INLET-GAS SPECIFICATIONS

The primary inlet flue gas specifications required are:

1. Volumetric flow rate.
2. Temperature.
3. Composition and molecular weight.
4. Amount of water vapor present, or the absolute humidity of the flue gas, or the adiabatic saturation temperature.
5. SO_2 concentration.

Also important is the particulate-phase concentration/size distribution in the inlet flue gas and its available alkalinity. However, the dry-particulate phase with its alkaline content, entering the spray chamber along with the flue gas, is not considered to be effective in SO_2 removal in the spray chamber. Its contribution primarily comes when recycled in the slurry along with fresh lime sorbent or other reagent, and is taken into account in the present model by specifying an appropriate recycle ratio and the active sorbent content in the recycled solids.

6.2 OPERATING PARAMETERS

There are basically five primary operating parameters which influence the bulk-gas properties at the spray-dryer outlet and SO_2 removal in the spray chamber:

1. Approach to saturation temperature at the spray-dryer outlet.
2. Stoichiometric ratio of fresh lime (sorbent) added in the atomized slurry to the amount of SO_2 in the inlet flue gas.
3. Recycle ratio of solids collected in spray dryer and particulate collection equipment in the slurry.
4. Slurry droplet diameter--method of atomization.
5. Method of slurry preparation (slaking)--primary lime particle size.

In the spray chamber, the gas is cooled and humidified adiabatically by evaporation of the water from the droplets. The temperature and absolute humidity of the gas phase at the spray-dryer outlet is then given by following the adiabatic saturation line on the psychrometric chart, from the flue-gas inlet conditions (see Figure 5). (Flue gas properties may be approximated to that of air to obtain a simple relationship for adiabatic saturation line due to Lewis Number = 1.) Thus, prescribing the approach to saturation temperature, ΔT, at the spray-dryer outlet sets the temperature and absolute humidity of the flue gas at the spray-dryer outlet.

The amount of moisture retained by the solids at the spray-dryer outlet is negligible compared to the amount of water evaporated. Thus, the amount of water added in the slurry per unit amount of inlet flue gas is simply obtained from the gas-phase humidities at the spray-dryer inlet and outlet.

The stoichiometric ratio in spray-dryer FGD operations is usually defined as moles of fresh lime or sorbent added in the slurry per mole of SO_2 in the inlet flue gas. Thus, setting the stoichiometric ratio determines the amount of fresh lime/sorbent added in feed slurry per unit amount of inlet flue gas. The recycle ratio further determines the amount of recycled solids (from particulate collection/spray dryer) accompanying the fresh lime in the slurry. Thus, the first three parameters fix the amount

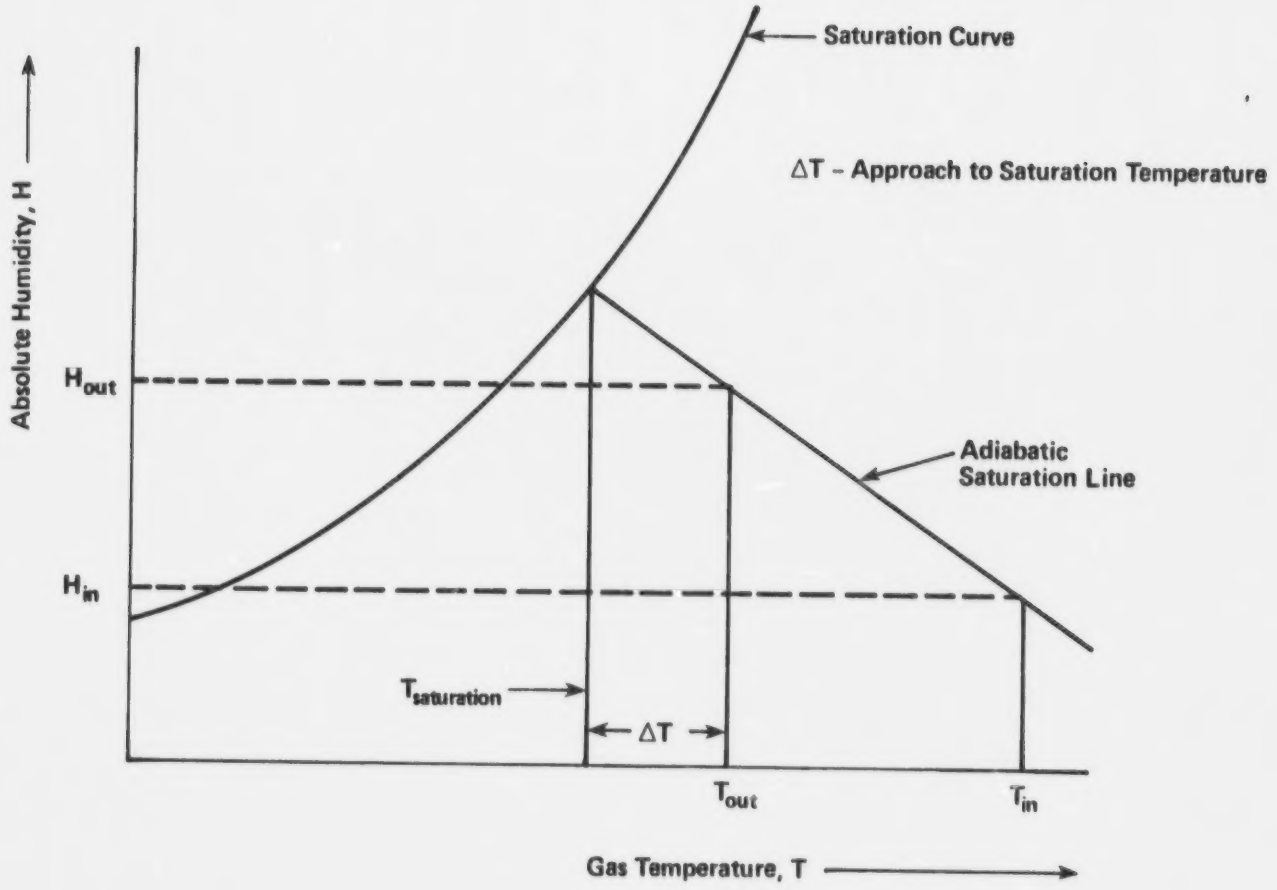


Figure 5. Adiabatic humidification of flue gas.

of total solids and water added in the slurry per unit amount of inlet flue gas, thus setting the solids concentration in the slurry.

The methods used for lime slaking, grinding of recycled solids, and slurry preparation determine the average primary particle size in the slurry. The recycle ratio and the amount of active reagent in the waste solids influence the active available alkaline content in the primary lime particles in the slurry. This in turn determines the equilibrium concentration of lime in the solution. The method of atomization, slurry concentration, and slurry feed rate determine the size distribution and average size of the atomized droplets as well as total number of droplets generated per unit amount of flue gas.

The amount of SO_2 that remains in the flue gas is not really an operating parameter; rather, it is the result of the other operating parameters. It is a design parameter, and the purpose of this modeling and experimental pilot-plant study is to predict it as a function of various operating variables.

6.3 GAS-FLOW PATTERN AND MIXING IN SPRAY CHAMBER

The inlet and operating conditions specify the bulk-gas properties at the inlet and outlet. However, the rates of heat and mass transfer depend upon the local properties of the bulk gas in contact with the spray droplet. The local bulk-gas properties (temperature and SO_2 and water-vapor concentrations) are determined by the gas-flow pattern and mixing within the spray chamber. With the completely backmixed gas-flow pattern, the bulk-gas properties throughout the spray chamber are uniform and same as at the spray-dryer outlet. At the other extreme, when the spray dryer is operated as an ideal plug-flow system, the bulk-gas properties change gradually with the local rate of change of a property depending upon the local rate of transfer (Levenspiel, 1962).

The plug-flow system generally gives greater transfer rates compared with backmixed systems. In a spray-dryer application, this means that the droplets will dry out faster in a plug-flow system as opposed to a backmixed system. This indicates a shorter time span for the wet-droplet stage in plug-flow spray-dryer application. The initial rate of SO_2 removal in a wet-droplet stage in a plug-flow system would be higher than that in a

backmixed system but will be operative for a shorter time period. Thus for SO_2 removal, both types of systems have their own strong points over the other. Plug-flow systems offer higher initial SO_2 removal rates, whereas backmixed systems offer longer drying times.

The actual gas-flow pattern and mixing would be intermediate of the above two extremes. Usually a deliberate attempt is made for efficient gas mixing in a spray dryer by employing efficient gas-distribution systems, such as a dual gas-dispersion system. In such situations, the gas mixing is more closely approximated by the well backmixed case, with uniform gas-phase temperatures and compositions in the spray chamber.

SECTION 7

MASS AND ENERGY BALANCES

Evaporation of water from droplets, absorption, and reaction of SO_2 in droplets change the bulk properties and composition of both the gas phase and droplets. The rate relations developed in Sections 4 and 5 must be coupled with material and energy balances for both the gas phase and droplets to develop the differential equations to describe the rate of change in the bulk properties (composition, mass, temperature, droplet diameter, etc.). These differential equations then must be integrated over the spray-dryer residence times of gas phase and droplets to obtain the overall removal of SO_2 in the spray chamber.

First, overall material and energy balances for the gas phase in the spray dryer need to be considered to establish all the initial parameters/properties of the droplet and gas phase. The total amount of water and the sorbent added in the spray chamber depend upon the operating parameters which include the approach to saturation at the spray-dryer outlet. The atomization parameters determine the average droplet diameter and drop-size distribution. The total number of droplets generated is calculated from the drop-size distribution and the total amount of sorbent slurry/solution added. After establishing the initial conditions, the gas phase and the droplets are "followed" from the spray-dryer inlet to outlet to determine total change in both by the above-mentioned differential equations.

Since the rigorous two-dimensional momentum balance equation is not considered in the present model, it is enough to consider one-dimensional flow of droplets from top to bottom of spray dryer. The droplets may be assumed to be uniformly distributed across the spray chamber cross section. The spray droplets may also be assumed to have the same residence time and mean velocity as that given by the gas-phase volumetric flow rate and mean velocity. In case of plug-flow type gas-flow pattern, the gas phase is also assumed to travel cocurrently with the droplets from the top to bottom of the spray dryer. In case of backmixed-flow type gas-flow pattern, the

gas phase is assumed to be uniformly mixed throughout the spray chamber and its properties are assumed to be those at the outlet. The spray droplets in this situation are exposed to constant gas atmosphere throughout the residence time.

In the following sections, the relationships for the overall mass and energy balances for the gas phase are first developed. The differential equations, describing the rate of change in droplet properties, are developed next and followed by those for the gas phase assuming a plug-flow type gas-flow pattern. No such relations are necessary for the backmixed type gas-flow pattern.

7.1 OVERALL GAS-PHASE MASS AND ENERGY BALANCE*

The inlet gas parameters, as described in Section 6, specify the inlet temperature, TI, °K; volumetric gas-flow rate, V, m^3/sec; and inlet SO_2 concentration, SI, mole fraction. Also, usually specified is the dry-gas composition and resulting molecular weight, MW, g/gmole. If not specified, the dry gas is assumed to be standard air with MW = 28.9.

In addition, the amount of water vapor present in the flue gas is needed, HI, g of water/g of dry air. This quantity may be specified in several ways: as above, as mole fraction of water, WI, mole fraction; adiabatic saturation temperature which also equals the wet-bulb temperature for an air-water system, TS, °K; or the dew point. The dew point can be easily converted into equivalent water mole fraction, WI, by using vapor pressure of water at the dew point. The relationships to convert WI to HI and TS to HI may be given as follows (assuming atmospheric pressure conditions):

$$HI = \frac{18.0}{MW} \frac{WI}{(1.0 - WI)} , \tag{32}$$

$$HS - HI = \frac{CS}{L} (TI - TS) , \tag{33}$$

*Symbols used in Sections 7 and 8 correspond to those used in the SPRAYMOD computer program.

where HS is the saturation humidity at the adiabatic saturation temperature, TS; L is the latent heat of water vaporization at TI, cal/g; and CS is the humid heat of air at TI, cal/g of dry air, °K, and for air-water is given by:

$$CS = 0.24 + 0.446 \times HI \ . \qquad (34)$$

Using the above relations, HI, TI, HS, and TS are obtained. Equation (33) assumes adiabatic saturation of the gas phase in the spray dryer. Thus, any heat losses in the spray-dryer system from the dryer walls are ignored, which provide cooling of the gas phase in addition to that by the spray droplets. At the adiabatic saturation temperature of the gas phase it is easier to calculate the inlet humidity of the gas phase using Equation (33); however, to calculate adiabatic saturation temperature knowing the inlet humidity requires trial and error method using Equation (33), since both HS and TS are interrelated and unknown.

The operating parameters for the spray dryer specify the approach to saturation at the spray dryer outlet, AT, °K. The gas-phase spray dryer outlet temperature, TO, is thus given by TO = AT + TS. The outlet gas-phase humidity, HO, assuming adiabatic saturation of the gas phase is given by:

$$HS - HO = \left(\frac{CS}{L}\right) \times (TO - TS) \ , \qquad (33')$$

where L is the latent heat of vaporization at TO, and CS is the humid heat of air at TO and is given by Equation (34) when HO replaces HI.

Similarly, the mole fraction of water in the outlet gas, WO, is given by Equation (32) when the outlet gas-phase properties supplant those for the inlet gas.

The fractional relative humidity of the outlet flue gas, RH, is given by:

$$RH = WO \times \frac{760}{PS} \ , \qquad (35)$$

where PS is the saturation vapor pressure at TO in mm of Hg.

The difference in the gas-phase humidities at the outlet and inlet (HO - HI) determine the amount of water evaporated in the spray chamber per

gm of dry-inlet gas. The amount of water retained by the dried particles at the outlet cannot be determined at this stage, but is usually very small and may be assumed to be 1 percent of the amount of water evaporated. Thus, the amount of water added in the spray chamber per gm of the inlet-dry gas, GW g of water in slurry/g of dry-inlet gas, is given by:

$$GW = 1.01 \times (HO - HI) \ . \tag{36}$$

The amount of fresh sorbent added is determined by the specified stoichiometric ratio, SR, gmole of sorbent/gmole of SO_2 in the flue gas. The recycle ratio, RR, and the amount of active sorbent in the recycled solids, as well as fresh solids, determine the mass fraction of active sorbent in the solids added in the slurry, LX. For pure sorbent and no recycle, LX is equal to 1. The amount (gm) of solids added in the slurry per gm of dry- inlet gas, GS, is then obtained as:

$$GS = \frac{SI}{(1.0 - WI)} \frac{SR\ MWS}{MW} (1 + RR) \ , \tag{37}$$

where MWS is the molecular weight of active sorbent. The amount of active sorbent in the slurry solids per gm of dry-inlet gas is given by (GS × LX).

Usually all the water and the sorbent solids are mixed together before atomization. The mass fraction of water in the feed slurry, XF is then given by:

$$XF = \frac{GW}{(GW + GS)} \ . \tag{38}$$

If only part of water is added with the slurry and the rest of it added separately, then the amount of water added with the slurry must be specified. This would also generate two sets of droplets with different compositions. Only the sorbent droplets will contribute to SO_2 removal, but all the droplets will contribute to gas humidification. For the present discussion, it will be assumed that all the water is added together and that the water mass fraction in the slurry is given by Equation (38).

The atomization parameters determine the droplet size distribution and a mean droplet diameter. Ideally, such size distribution should be used,

but often this information is not available or measured. Therefore, for model simplicity, it will be assumed that all droplets are of the same initial diameter, DO. The density of the slurry/solution may be estimated from mass fraction, XF, sorbent solid's density, ρ_s, and taking density of water as 1 g/cm^3. The initial weight of a droplet, WO, may then be obtained by:

$$WO = \frac{\pi}{6} DO^3 \frac{\rho_s}{1 + (\rho_s - 1)\ XF} . \tag{39}$$

The number of slurry droplets atomized per gm of dry-inlet gas, TN, may then be obtained by:

$$TN = \frac{GW}{WO\ XF} . \tag{40}$$

When the sorbent is injected in the form of slurry, the slaking and grinding processes determine the sorbent particle size distribution and mean size. The size distribution of these particles is also often not available. A mean sorbent particle diameter, DP, may be assumed to represent the distribution. To determine the dissolution rate of sorbent in a droplet, the total number of primary particles in a droplet, NP, is required and is given by:

$$NP = \frac{DO^3\ (1 - XF)}{DP^3\ [1 + (\rho_s - 1)\ XF]} . \tag{41}$$

Finally, the residence time of the droplets and the gas phase in the spray dryer, RT, sec, is given by the volumetric gas-flow rate, V, and the volume of the spray dryer, VS, as:

$$RT = \frac{VS}{V} . \tag{42}$$

The above equations and spray-dryer initial and operating parameters determine the necessary initial conditions for both the gas phase and the droplets.

7.2 DROPLET MASS AND ENERGY-BALANCE EQUATIONS

The evaporation of water changes the moisture content in the droplet, and the absorption/reaction of SO_2 changes the composition of solids and reduces the amount of available sorbent. The heat transfer from the gas phase provides the heat of vaporization and also affects the droplet temperature. In the initial period of constant drying, the droplet temperature remains at the gas-phase adiabatic saturation temperature, and it begins to rise as the amount of moisture falls below the critical moisture content.

The initial (t = 0) weight of a droplet, WO, is given by Equation (39). The initial weight of water in a droplet, WW, is given by (XF × WO). The initial amount of active sorbent in a droplet, CM, is given by:

$$CM = WO\ (1 - XF)\ LX\ . \tag{43}$$

The initial droplet temperature, TW, is specified by sorbent-feed conditions. At a given time (t = t), the rates of change in weight of moisture in a droplet, W; the weight of active sorbent in a droplet, S; and the droplet temperature, TD, are related to the rate of mass and heat transfer between the droplet and the gas phase, and are given as:

$$\frac{dW}{dt} = -\ 18\ N_W\ , \tag{44}$$

$$\frac{dS}{dt} = -\ MWS \times N_S\ , \tag{45}$$

$$\frac{\pi}{6} D^3\ \rho\ C_p \frac{dTD}{dt} = Q - N_W\ L\ , \tag{46}$$

where N_W is the molar rate of water evaporation as given by Equation (7) or (11) or (14);

N_S is the molar rate of SO_2 transfer from gas phase to droplet as given by Equation (15) or (17) or (23);

Q is the rate of heat transfer from gas phase to the droplet as given by Equation (8) or (12);

D is the droplet diameter at time t; and

L is the latent heat of vaporization at TD.

Note that for evaporation, N_W would be positive; hence, the negative sign in Equation (44). Similarly, N_S is positive and hence the negative sign in Equation (45), and Q and N_W are both positive in Equation (46). In Equation (45) equimolal reaction is assumed between the sorbent and SO_2. The above relations describe the mass and energy balances around a droplet. The change in the total weight of a droplet, WD, is given by combining the loss of water and the addition of SO_2:

$$\frac{d\ WD}{dt} = -\ 18\ N_W + 64\ N_S\ . \tag{47}$$

The moisture content in the droplet, X, is given by:

$$X = \frac{W}{WD}\ . \tag{48}$$

If the density of the product solids and that of the initial solids is assumed to be the same (ρ_S), then the droplet diameter, D, at a time, t, with moisture-mass fraction in the droplet, X, and total droplet weight, WD, is given by:

$$D = \left[\frac{6}{\pi}\ \frac{WD + (\rho_S - 1)\ W}{\rho_S}\right]^{1/3}\ . \tag{49}$$

When the moisture content, X, falls below a critical moisture content, X_c, the diameter of the droplet is assumed to remain constant during further drying.

Equations (44)-(49) describe the changes in the moisture and sorbent content of a droplet with time. These equations may be integrated over

time for the residence time of the droplets, RT, using the initial conditions mentioned above. The utilization of the sorbent, UT, is given by the remaining amount of active sorbent, S, and the initial amount of sorbent, CM:

$$UT = \frac{(CM - S)}{CM} . \tag{50}$$

The efficiency of SO_2 removal, EF, is then given by multiplying the stoichiometric ratio of the active sorbent available, SR, with the sorbent utilization, UT:

$$EF = UT \times SR . \tag{51}$$

To solve the above equations, all the parameters required to determine the rate relations N_W, N_S, and Q must be specified. The parameters are described in Sections 4 and 5 and include critical moisture content, equilibrium moisture content, gas-phase diffusivity values, gas-phase thermal conductivity, equilibrium sorbent solubility, solid-phase reaction coefficient, etc.

7.3 GAS-PHASE MASS- AND ENERGY-BALANCE EQUATIONS

To determine the heat- and mass-transfer rates to and from droplets, it is necessary to determine the gas-phase properties (temperature and concentrations of SO_2 and water) throughout the spray chamber.

i) Backmixed flow pattern--this task is simpler when the gas flow is assumed to be completely mixed, when the gas composition and temperature are the same as at the outlet. The gas-phase temperature and water vapor concentration at the spray-dryer outlet, TO and WO, are determined by the overall balance as discussed before. The outlet SO_2 concentration is however not known a priori. In this case a trial-and-error method must be used to determine the outlet SO_2 concentration SO, and hence the removal efficiency of the spray-dryer system. A trial value of SO is first assumed. The droplet Equations (44)-(49) are then solved starting from initial conditions until the residence time is reached. The SO_2 removal efficiency,

EF, is then calculated using Equation (51). The model predicted value of SO is then calculated as:

$$SO = (1 - EF)\ SI\ . \tag{52}$$

The model predicted value of SO is then compared with the assumed value of SO. Using the difference, a new value of SO is assumed. The process is repeated until the correct value of SO is found.

Thus in this case, since the gas-phase concentrations are constant throughout the spray chamber, no gas-phase balance equations are required.

ii) Plug-flow pattern--if the gas flow in the spray chamber is assumed to be plug-flow type, the gas-phase concentrations and temperature change gradually from top to bottom of the spray chamber. The gas flow is further assumed to be uniform across the spray chamber cross section with constant properties across a section. This is a somewhat idealized and unrealistic situation for a spray chamber.

The changes in gas-phase concentration and temperature are related with the rates of water evaporation, SO_2 absorption, and heat transfer. Since the spray droplets may be assumed to be uniformly mixed throughout the gas phase, it is sufficient to "follow" a small parcel of gas phase with 1 gm of dry inlet gas along with, TN, sorbent droplets, moving together for a period of gas-phase residence time, RT.

The initial conditions (t = 0) are given by those at the inlet. The gas temperature at inlet is TI, water vapor mole fraction is WI, and SO_2 mole fraction at inlet is SI. At a given time t after gas entry (t = t), the rates of change in the gas-phase water vapor mole fraction, XW; SO_2 mole fraction, XS; and the gas temperature, TG, are given by:

$$\frac{1}{MW\ (1 - XW)} \frac{d\ XW}{dt} = N_W\ TN\ , \tag{53}$$

$$\frac{1}{MW\ (1 - XW)} \frac{d\ XS}{dt} = -\ N_S\ TN\ , \tag{54}$$

$$CS\ \frac{d\ TG}{dt} = -\ Q\ TN\ , \tag{55}$$

where N_W is the molar rate of water evaporation as given by Equations (7) or (11) or (14);

N_S is the molar rate of SO_2 transfer from gas phase to droplet as given by Equations (15) or (17) or (23);

Q is the rate of heat transfer from gas phase to droplet as given by Equations (8) or (12);

TN is the number of droplets per gm of dry-inlet gas;

MW is the molecular weight of dry-inlet gas; and

CS is the humid heat of the gas of dry-inlet gas.

The humid heat, CS, is given by:

$$CS = 0.24 + 0.446 \frac{18\ XW}{MW\ (1 - XW)} . \qquad (56)$$

Equations (53)-(55) describe the changes in gas-phase composition and temperature with time. With the initial conditions mentioned above, these equations may be integrated numerically over the gas-phase residence time, RT. Like the droplet equations, the SO_2 removal efficiency may also be determined from the gas-phase SO_2 concentration at the outlet. If the SO_2 concentration at the spray-dryer outlet is calculated to be SO, the removal efficiency, EF, may be obtained using Equation (52). To solve the above equations, all the parameters required to determine the rate relations N_W, N_S, and Q must be specified.

SECTION 8

"SPRAYMOD" COMPUTER PROGRAM

To solve the material and energy balance equations described in the last section using the rate relations developed in Sections 4 and 5, a computer program, "SPRAYMOD," was written in Basic language. The computer program was developed for a desk-top microcomputer.

The program basically has three sections. In the first section, a menu input format is used to enter all the input data regarding specifications and operating variables. After input, the data are printed for verification. In the second section, the gas-phase overall mass and energy balances are carried out to establish all the initial conditions for both the droplet and the gas phase. All the dependent variables are then initialized.

In the third section, the differential equations for the droplets and for the gas phase (plug-flow option) are solved by a simple, explicit, finite-forward-difference scheme. The time step is controlled so that a maximum change during a time step in the droplet temperature or the droplet weight is less than 1 percent of the function value. This criterion ensures accurate solution in spite of the simple numerical scheme used for the solution.

The program has two main options regarding the gas-flow pattern: 1) backmixed flow and 2) plug flow. The backmixed-flow case requires a trial-and-error procedure to determine SO_2 removal efficiency. The backmixed flow assumption is close to the real situation. Since the backmixed case needs a first trial value for SO_2 removal efficiency, this value may be obtained by first running the program with the plug-flow option.

The program essentially follows a single spray droplet suspended in the bulk-gas phase with time. The rates of mass and heat transfer are related to changes in droplet and bulk-gas properties (e.g., droplet diameter and gas-phase SO_2 concentration). The bulk-gas physical properties required in various rate relations are evaluated at the mean temperature, TM, of the gas film surrounding the droplet, TM = (TG + TD)/2. The

properties of the liquid phase, such as saturation vapor pressure, are evaluated at the droplet temperature, TD, which is assumed to be uniform throughout the droplet.

At initial time, t = 0, all the bulk gas and droplet properties are initialized. If the backmixed flow option is used, the first trial value of SO_2 removal efficiency needs to be specified. At each time increment, the derivative functions are determined and the time step, DT, is established. The amount of water evaporated and SO_2 absorbed in time DT are then evaluated. Corresponding changes in all gas-phase and droplet variables during time step, DT, are determined. The calculations continue until no more SO_2 is removed by droplets or until the time reaches the residence time value.

Figure 6 shows a schematic flow chart of the program SPRAYMOD. The complete listing of this Basic program is given in Appendix A. This program uses several equations to determine physical properties of both the gas and liquid phases. Thermal conductivity and gas-phase diffusivities are assumed to be those with respect to air.

1. Latent heat of evaporation of water, L, cal/gm, at a temperature, T, °K:

$$L = 597.64 - 0.55\ (T - 273.15)\ .$$

2. Vapor pressure of water, PS, mm of Hg, at a temperature, T, °K:

$$PS = EXP\ (20.426 - 5138.8/T)\ .$$

3. Diffusivity of SO_2 in air, DS, cm^2/sec, at temperature, T, °K:

$$DS = 2.951\times10^{-5}\times T^{1.5}\ .$$

4. Diffusivity of water in air, DW, cm^2/sec, at temperature, T, °K:

$$DW = 5.062\times10^{-5}\times T^{1.5}\ .$$

5. Molar density of gas phase, CT, $gmole/cm^3$, at temperature, T, °K:

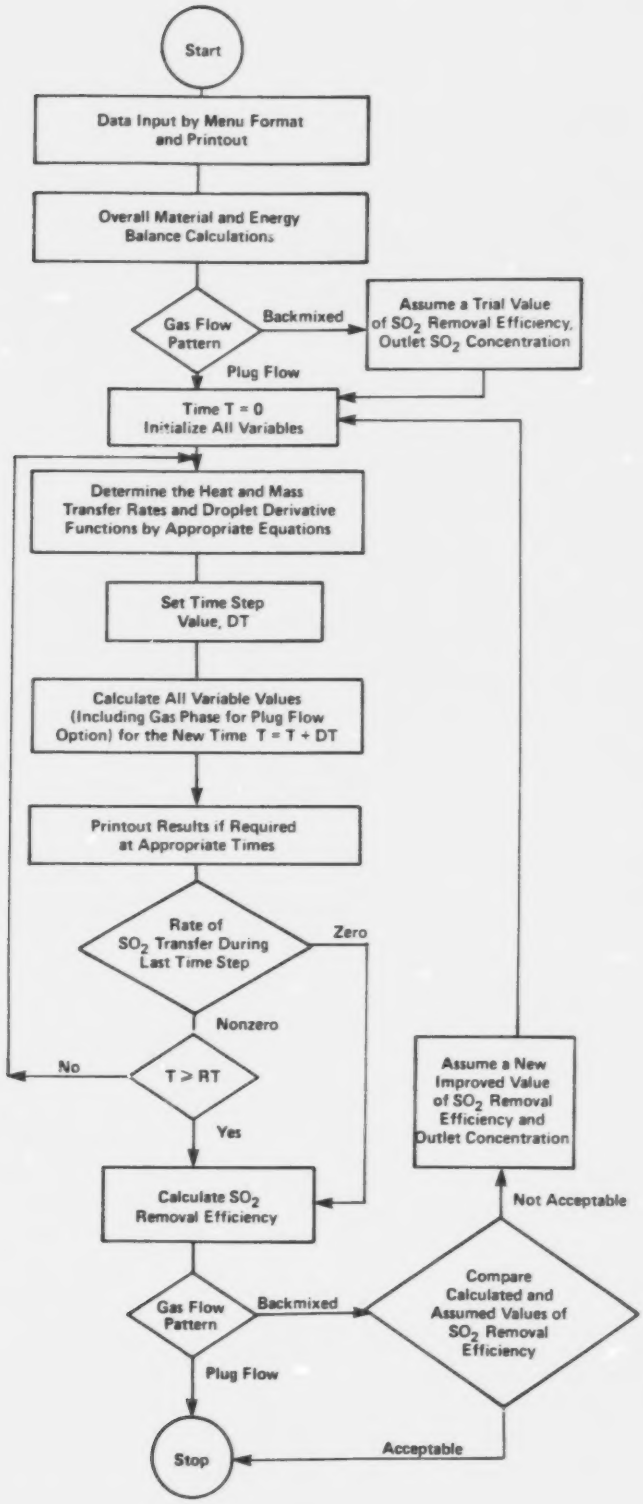

Start
Data Input by Menu Format and Printout
Overall Material and Energy Balance Calculations
Gas Flow Pattern
Backmixed
Assume a Trial Value of SO_2 Removal Efficiency, Outlet SO_2 Concentration
Plug Flow
Time T = 0 Initialize All Variables
Determine the Heat and Mass Transfer Rates and Droplet Derivative Functions by Appropriate Equations
Set Time Step Value, DT
Calculate All Variable Values (Including Gas Phase for Plug Flow Option) for the New Time T = T + DT
Printout Results if Required at Appropriate Times
Rate of SO_2 Transfer During Last Time Step
Zero
Nonzero
No
T > RT
Yes
Assume a New Improved Value of SO_2 Removal Efficiency and Outlet Concentration
Not Acceptable
Calculate SO_2 Removal Efficiency
Gas Flow Pattern
Backmixed
Compare Calculated and Assumed Values of SO_2 Removal Efficiency
Plug Flow
Stop
Acceptable

Figure 6. Flowchart for program SPRAYMOD.

$$CT = 1.2194 \times 10^{-2}/T \ .$$

6. Thermal conductivity of air, CO, cal/cm-sec-°C, at temperature, T, °K:

$$CO = 1.82 \times 10^{-7} \ T + 7.9 \times 10^{-6} \ .$$

7. Diffusivity of $Ca(OH)_2$ in water, DC, cm^2/sec, at T, °K:

$$DC = 6.133 \times 10^{-8} \times T \ .$$

8. Equilibrium solubility of $Ca(OH)_2$ in water, CL, gm/cm^3, at temperature, T, °K:

$$CL = 1.88 \times 10^{-3} - 1.174 \times 10^{-5} \ (T - 273.15) \ .$$

8.1 PROGRAM SIMULATION

Figure 7 shows the results of a sample simulation under assumed operating conditions. This basically shows operability of the model. Several parameters were assumed to carry out the model calculations; thus, this sample represents a hypothetical case. The key parameters and operating variables used in this particular model simulation are given below.

Operating Variables:

- Inlet gas temperature = 160°C
- Amount of water in inlet flue gas = 6.4% by volume
- SO_2 concentration at inlet = 800 ppm
- Approach to saturation temperature at the spray dryer outlet = 15°C
- Average droplet diameter = 50 μm
- Primary lime particle size = 4 μm
- Stoichiometric ratio = 1.0
- Recycle ratio = 0

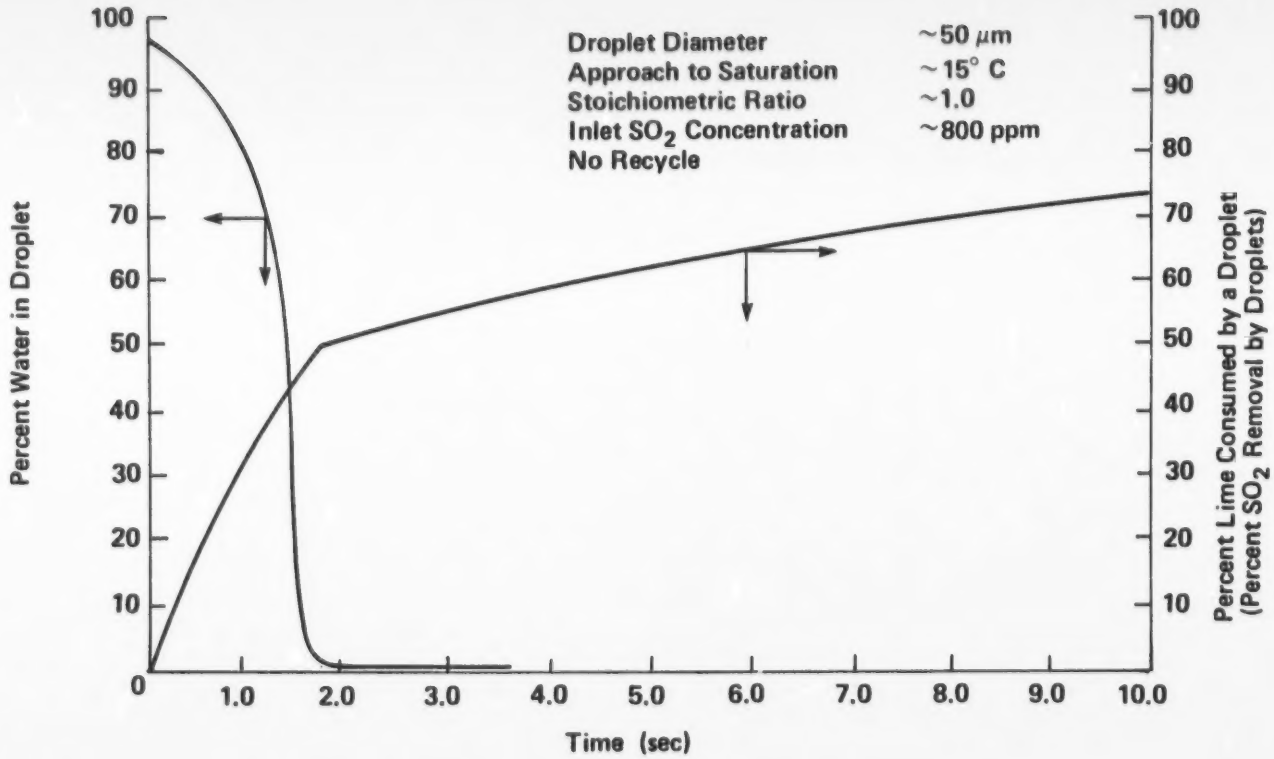


Figure 7. Evaporation of water and SO_2 absorption by a droplet in a spray dryer.

- Mass fraction of active sorbent in feed = 1.0

Key Parameters Assumed:

- Critical moisture content = 50%
- Equilibrium moisture content at 100% relative humidity = 3%
- Completely backmixed flow system in spray dryer
- Volumetric reaction coefficient at 2% residual moisture in solids = 1×10^7 cm^3/gmole-sec

The simulation shows that, in this particular case, the slurry droplet loses most of its moisture in the first couple of seconds with the moisture content eventually approaching its equilibrium value. Most of the SO_2 is absorbed in the first stage of drying with SO_2 removal continuing at a slower rate in the dry stage.

For the operating conditions and key parameters assumed, the SO_2 concentration at the spray dryer outlet was determined to be $\approx$ 215 ppm by a trial-and-error method. This was equivalent to 73 percent SO_2 removal efficiency in the spray dryer.

8.2 COMPARISON OF MODEL PREDICTIONS WITH PILOT-PLANT DATA

The predictions of SO_2 removal efficiency of a spray-dryer system by the model presented in this paper were compared with a set of pilot-plant data under various operating conditions. The pilot-plant data set used for comparison here was collected by Cottrell Environmental Sciences, Inc., at the Comanche Station of Public Service Company of Colorado (Stevens et al., 1984). One of the difficulties in obtaining a suitable data set for comparison was lack of complete information in various reported pilot-plant and full-scale studies. This particular data set was chosen because of its availability and extent of information; these pilot-plant studies were recently published by Stevens et al. (1984).

In addition to various operating parameters, some physical parameters need to be specified in model simulations. The drying characteristics of solids used were: critical-moisture content of 30% by mass and an

equilibrium-moisture content of 15% by mass at 100% relative gas-phase humidity. The critical-moisture content value was considered to be at which the solid spheres start touching each other in the droplet, whereas the equilibrium-moisture content value was approximately determined from the measured moisture contents of spray-dryer solids in this pilot-plant study (Stevens et al., 1984). The information regarding reaction coefficients was not available in this report. For model simulations, the solid-phase reactivity was therefore considered to be zero. No measured atomized droplet-size distribution was available in this study, and the mean-droplet size inferred from the measured spray-dryer outlet solid-size distribution was taken as 50 μm (monodispersed). The lime particle-size distribution in slurry was available and was approximately monodispersed with a mean size of 4 μm. The gas-flow pattern in model simulations was assumed to be completely backmixed.

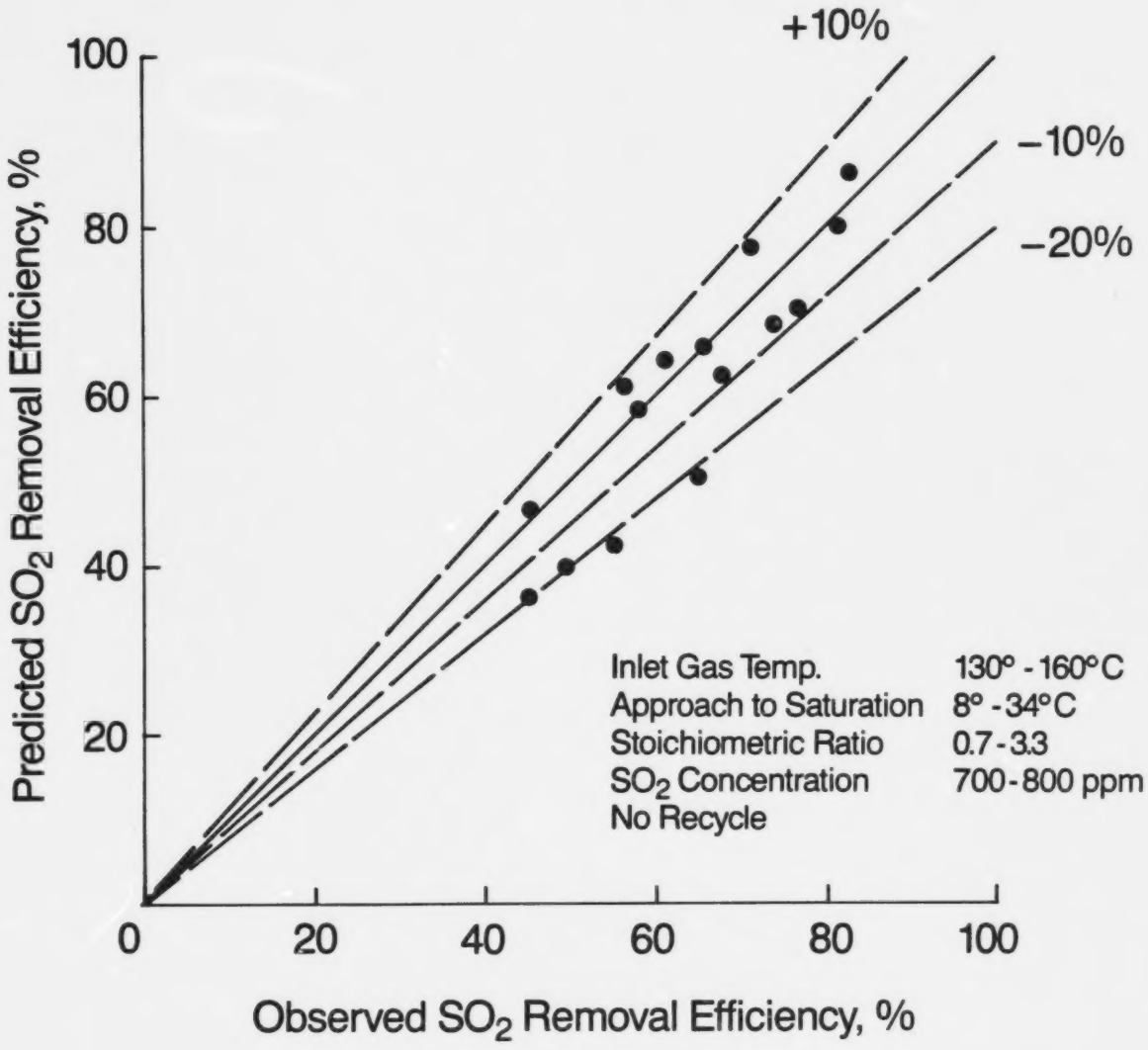
The model comparison with the data is shown in Table 1 and Figure 8. The agreement was very good, with most of the model prediction being within ±10% of data values. With closer approach to saturation, however, there was an underprediction of SO_2 removal efficiency. This is believed to be due to neglecting the solid-phase reactivity, which is expected to be significant at closer approach to saturation, due to increased amount of equilibrium moisture in solids.

TABLE 1. COMPARISON OF SO_2 REMOVAL EFFICIENCIES PREDICTED BY MODEL TO THOSE OBSERVED IN RESEARCH COTTRELL PILOT-PLANT DATA

Run Number	Inlet Gas Temperature (°F)	Adiabatic Saturation Temperature (°F)	Approach to Saturation (°F)	Stoichiometric Ratio	Inlet SO_2 Concentration (ppm)	Observed SO_2 Removal Efficiency (%)	SO_2 Removal Efficiency Predicted by Model (%)
101	334	125	69	3.26	780	57.7	58.6
104	340	125	33	1.69	710	76.8	70.4
105	301	126	74	2.36	780	44.9	46.4
107	300	126	52	2.70	730	56.2	61.4
109	307	126	62	1.28	800	45.0	36.3
111	301	126	34	1.62	780	65.4	65.4
112	307	126	22	1.55	750	81.3	79.8
115	256	126	31	1.56	770	61.0	64.0
116	261	127	15	1.30	800	71.2	77.5
118	340	125	15	1.31	810	82.7	86.4
120	341	124	61	1.35	790	49.4	39.2
125	262	126	18	0.71	800	65.0	50.0
126	264	126	18	0.99	800	67.5	62.6
130	340	125	20	1.17	800	73.8	68.8
131	342	123	22	0.72	800	55.0	42.0

Note: Recycle ratio = 0.

Reference for measured data - Stevens et al., 1984.



Source of observed efficiency data points—Stevens et al., 1984.

Figure 8. Comparison between predicted and observed SO_2 removal efficiencies under various operating conditions.

SECTION 9

SO_2 REMOVAL IN PARTICULATE CONTROL EQUIPMENT

As the spray-dried particles are carried away by flue gas and collected in a particulate collection device, they are likely to remove some additional SO_2. In the spray dryer, flue gas contacts only the sorbent fed into the spray dryer; but in a particulate collection device, the gas phase may contact solids accumulated over a long length of time. This effectively tends to provide a very high stoichiometric ratio in the particulate collection device. The active reagent in the dust deposited on the filter bags or on electrostatic precipitator walls provides the additional SO_2 removal.

Since the physical state of the particles in dust deposits is similar to that of the solids in the dry-particle stage in the spray chamber, related rate relations are applicable. The deposited dust is exposed continuously to the flue gas for much longer time periods as compared to spray-dryer residence time. The dust particles therefore may be assumed to be saturated with the SO_2 concentrations that they are exposed to. In other words, the diffusion process of SO_2 into the solid particles may not be important here due to the long exposure times. The rigorous analysis approach described in the preceding section is thus not required here. The amount of SO_2 removed would be entirely determined by the reactivity of the dust and the gas-phase SO_2 concentration at the dust surface. Thus, the simple approach of bulk-volume reactivity (K_r) would be applicable.

The reactivity, K_r, of the dust is again strongly dependent upon the moisture content of the solids. It would also be dust-specific and depend upon dust composition, type of reagent, average particle size, and available surface area. However, for a given source-reagent system, the reactivity would basically depend on the moisture content as the only variable process parameter. The equilibrium moisture content of solids is in turn dependent upon the relative humidity of the bulk-gas phase.

9.1 BAGHOUSE AS A PARTICULATE CONTROL DEVICE

As the flue gas passes through the dust cake, SO_2 is transported from the bulk-gas phase to the dust surface. Due to intimate dust-gas contact, the SO_2 concentration at the dust surface will be the same as the bulk gas-phase SO_2 concentration (Figure 9).

The SO_2 removal rate in the dust cake is given by:

$$V(C_{SO_2,in} - C_{SO_2,out}) = K_r \, C_{SO_2,mean} \, C_{unreacted\ lime} \, A \, t_{dust} \,, \qquad (57)$$

where K_r is the volumetric reaction coefficient; V is the volumetric gas flow rate; A is the surface area of fabric filter; t_{dust} is the dust layer thickness; $C_{SO_2,in}$ and $C_{SO_2,out}$ are molar gas-phase concentrations of SO_2 at the inlet and at the outlet of the fabric filter, respectively; $C_{unreacted\ lime}$ is the molar concentration of the available lime reagent in the dust cake solids; and $C_{SO_2,mean}$ is taken as the arithmetic mean of inlet and outlet SO_2 concentrations. Thus, efficiency of SO_2 removal, η, is given by:

$$\eta = \frac{C_{SO_2,in} - C_{SO_2,out}}{C_{SO_2,in}} = \frac{K'}{1 + K'/2} \,, \qquad (58)$$

where $K' = K_r \, C_{unreacted\ lime} \, \frac{A}{V} \, t_{dust}$.

The ratio V/A represents the air-to-cloth ratio for the fabric filter.

Thus, the efficiency of SO_2 removal in a fabric filter is expected to increase with higher reactivity (K_r), greater concentration of unreacted sorbent, lower air-to-cloth ratio, and thicker dust cake. The latter two parameters lead to longer gas-phase residence time in dust cake.

9.2 ELECTROSTATIC PRECIPITATOR AS A PARTICULATE CONTROL DEVICE

For SO_2 removal in an ESP, SO_2 has to be transported from the bulk-gas phase to dust deposited on collection plates by gas-phase diffusion. Therefore, the gas-phase diffusion resistance contributes to overall SO_2 removal in addition to dust reactivity coefficient, K_r.

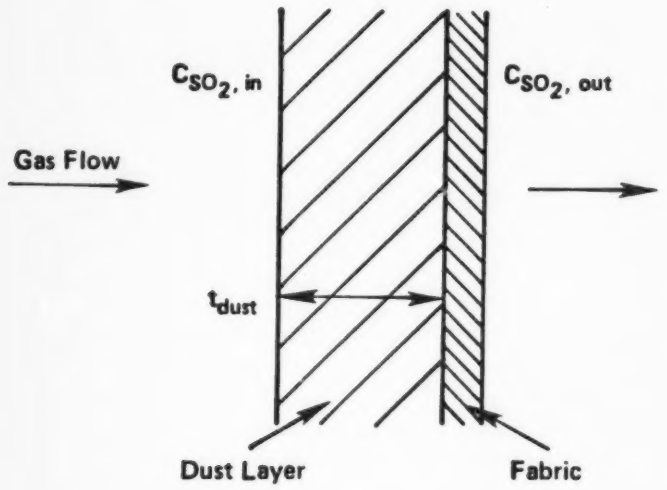


Figure 9. Schematic of SO_2 removal in a baghouse.

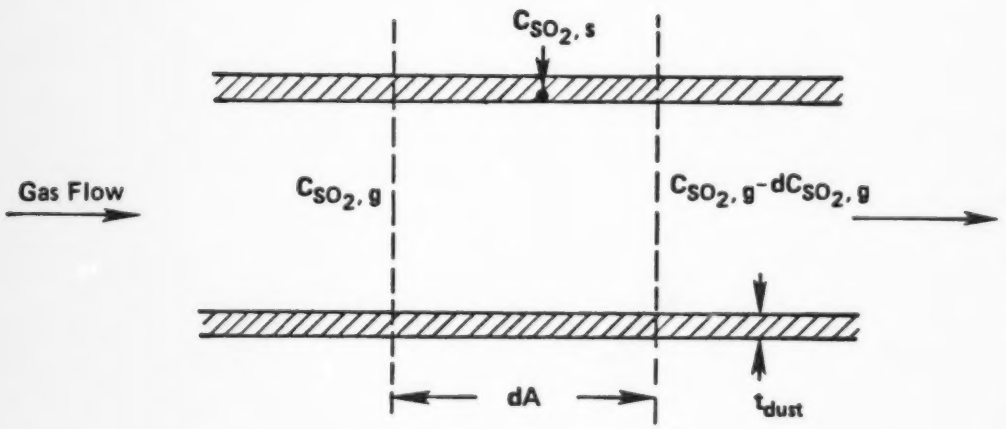


Figure 10. Schematic of SO_2 removal in an electrostatic precipitator.

A schematic of an ESP section is shown in Figure 10. By taking a mass balance on the small section shown in this figure:

$$\text{Rate of } SO_2 \text{ removal} = - V \, dC_{SO_2,g} \,, \tag{59}$$

$$= k_d \, (C_{SO_2,g} - C_{SO_2,S}) \, dA \,,$$

$$= K_r \, C_{SO_2,S} \, C_{\substack{\text{untreated} \\ \text{lime}}} \, t_{dust} \, dA \,,$$

where k_d is the gas-phase mass transfer coefficient; A is the surface area of collection plates; $C_{SO_2,g}$ is the concentration of SO_2 in the bulk-gas phase; $C_{SO_2,S}$ is the gas-phase concentration of SO_2 at the dust layer; K_r is the volumetric reaction coefficient; t_{dust} is the thickness of the dust layer deposited on collection plates; and V is the volumetric gas flow rate. Eliminating surface SO_2 concentration [$C_{SO_2,S}$ in Equation (59)]

$$- V \, dC_{SO_2,g} = k_d \, C_{SO_2,g} \, \phi \, dA \,, \tag{60}$$

where ϕ is given by:

$$\frac{\phi}{1 - \phi} = \frac{K_r \, C_{\text{unreacted lime}} \, t_{dust}}{k_d} \,. \tag{61}$$

ϕ is enhancement factor, a measure of relative contribution of reaction rate and diffusional transport. For a high reaction rate

$$K_r \, C_{\text{unreacted lime}} \, t_{dust} >> k_d \,,$$

and

$$\phi \text{ approaches } 1 \,.$$

ESPs generally have very well-defined flow geometry (flow through two parallel plates) and, hence, experimental mass transfer correlations reported in the literature for similar flow geometries may be used to estimate the mass transfer coefficient, k_d.

For example, one correlation for confined-flow through two parallel plates gives (Treybal, 1980):

$$k_d = 0.11\ U\ Re^{-0.29} Sc^{-0.67} , \qquad (62)$$

$$\text{for } Re \approx 2{,}500\text{-}25{,}000 ,$$

where U is the gas velocity through the parallel plates, Re is the gas flow Reynolds Number, and Sc is the Schmidt Number for SO_2-air system. Thus, the mass-transfer coefficient, k_d, may reasonably be estimated. The reaction coefficient, K_r, as discussed before is primarily dependent on the moisture content of dust.

For constant operating conditions, ϕ, as given by Equation (61), would be reasonably constant and Equation (60) may be integrated over the entire length of the precipitator to give the SO_2 removal efficiency of an ESP, η, as:

$$\eta = \frac{C_{SO_2,in} - C_{SO_2,out}}{C_{SO_2,in}} = 1\text{-}e^{-k_d\ \phi\ \frac{A}{V}} . \qquad (63)$$

The ratio (A/V) is the specific collection area (SCA) of an ESP. Higher efficiency of SO_2 removal in an ESP is expected for greater values of the mass-transfer coefficient, ϕ, and SCA. Higher values of enhancement factor, ϕ, are obtained with higher reactivity, K_r, greater concentration unreacted sorbent, thicker dust cake, and lower mass-transfer coefficient.

Equations (58) and (63) relate the SO_2 removal efficiency of particulate control equipment to various parameters such as K_r (reactivity and moisture content), concentration of unreacted lime in dust, dust layer thickness, mass-transfer coefficient k_d, air-to-cloth ratio for baghouses, and SCA for ESP. By analyzing data on baghouse/ESP pilot plants for SO_2

removal efficiencies, an estimation of the reaction coefficient (K_r) and its variation with moisture content may be determined.

Wilkinson and Tonn (1982) reported SO_2 removal efficiencies obtained in a Babcock & Wilcox dry FGD pilot plant, using both baghouse/ESP, under a variety of operating conditions. Since the amount of unreacted lime was not given in these studies, only the product, P, which is (K_r $C_{unreacted\ lime}$ t_{dust}), could be determined by analyzing these data. These values are given in Tables 2-5. Electrostatic precipitators generally tolerate a thicker dust layer as compared to a baghouse, and this is reflected in higher product, P, values for similar operating conditions; e.g., for spray dryer approach temperature of 30°F (16.67°C), control device approach temperature of 25°F (13.89°C), and stoichiometric ratio of 1.4, the product P for baghouse was about one-half that for an ESP.

More detailed data would enable estimating the reactivity coefficient separately.

From the modeling of SO_2 removal processes in baghouse and ESP, their relative effectiveness may be described as follows.

1) A baghouse does not offer any mass transfer resistance to SO_2 removal; whereas, this is a significant factor in ESP. Thus, for similar operating conditions such as moisture content of solids (same approach to saturation temperature), dust layer thickness, total surface area, and concentration of unreacted lime, a baghouse is expected to be more effective in SO_2 removal compared to an ESP.

2) ESPs can generally be operated at a temperature closer to saturation than can a baghouse. While operating a baghouse, one has to be concerned about not blinding the bags. This puts a lower limit to the approach temperature at which a baghouse may be operated. A closer approach temperature in ESP means higher moisture content in solids depositing on ESP walls. This would give a higher dust reactivity in an ESP compared to a baghouse.

3) Another factor in favor of ESP is that it can tolerate more dust buildup on walls than a baghouse can (due to pressure drop limitations on a baghouse). This allows accumulation of a greater amount of active reagent on the ESP walls which should contribute more to SO_2 removal.

TABLE 2. EFFECT OF APPROACH TO SATURATION TEMPERATURE AT THE BAGHOUSE OUTLET ON SO_2 REMOVAL EFFICIENCY IN A BAGHOUSE

Approach to Saturation (°F)	Efficiency (η)	P (cm/sec)
50	0.032	0.033
45	0.058	0.062
40	0.096	0.103
35	0.165	0.183
30	0.249	0.290
25	0.340	0.418
20	0.421	0.544

$P = K_r C_{\text{unreacted lime}} t_{\text{dust}}$.

Air-to-cloth Ratio = 1.0 cm/sec.

Stoichiometric Ratio = 1.0 at spray dryer inlet.

Approach to Saturation in Spray Absorber = 30°F.

TABLE 3. EFFECT OF STOICHIO RIC RATIO ON SO_2 REMOVAL EFFICIENCY IN A BAGHOUSE

Stoichiometric Ratio	Efficiency (η)	P (cm/sec)
0.6	0.229	0.264
0.8	0.28	0.332
1.0	0.334	0.409
1.2	0.379	0.477
1.4	0.422	0.546
1.6	0.463	0.614

Air-to-cloth Ratio = 1.0 cm/sec.

Approach to Saturation in Spray Absorber = 30°F.

Approach to Saturation in Baghouse = 25°F.

TABLE 4. EFFECT OF APPROACH TO SATURATION TEMPERATURE AT ESP OUTLET ON SO_2 REMOVAL EFFICIENCY IN AN ESP

Approach to Saturation (°F)	Efficiency (η)	P (cm/sec)
40	0.162	0.774
35	0.168	0.855
30	0.175	0.964
25	0.183	1.116
20	0.2	1.569

Mass Transfer Coefficient, k_d = 0.54 cm/sec.

Stoichiometric Ratio = 1.4 at inlet of spray dryer.

Temperature Drop in ESP = 5°F.

TABLE 5. EFFECT OF STOICHIOMETRIC RATIO ON SO_2 REMOVAL EFFICIENCY IN AN ESP

Stoichiometric Ratio	Efficiency (η)	P (cm/sec)
0.6	0.108	0.332
0.8	0.132	0.483
1.0	0.153	0.671
1.2	0.182	1.097
1.4	0.2	1.569

Mass Transfer Coefficient, k_d = 0.54 cm/sec.

Approach to Saturation in Spray Absorber = 25°F.

Approach to Saturation in ESP = 20°F.

SECTION 10

NOMENCLATURE

A - surface area, cm^2

c - molar density of gas phase, $gmoles/cm^3$

C - concentration, $gmoles/cm^3$

C_p - specific heat (gas phase), cal/gm °C

$C_{p,l}$ - specific heat (liquid phase), cal/gm °C

d_d - droplet diameter, cm

d_p - lime particle diameter, cm

D - diffusivity, cm^2/sec

g - gravitational acceleration, cm/sec^2

h - gas-phase heat transfer coefficient, $cal/cm^2\text{-}sec\text{-}°C$

H - Henry's Law constant, $\frac{gmole/cm^3 \text{ in gas phase}}{gmole/cm^3 \text{ in liquid phase}}$

k - gas-phase thermal conductivity, cal/cm-sec-°C

k_d - gas-phase mass transfer coefficient, cm/sec

k_l - liquid-phase mass transfer coefficient, cm/sec

k_o - overall mass transfer coefficient, cm/sec

K_r - volumetric reaction coefficient, cm^3/sec-gmole of sorbent

N_s - molar rate of gas-phase mass transfer of SO_2, gmole/sec

Nu - Nusselt Number, dimensionless

N_w - molar rate of gas-phase mass transfer of water, gmole/sec

Pr - Prandtl Number, dimensionless

Q - rate of gas-phase heat transfer, cal/sec

r - radial coordinate, cm

Re - Reynolds Number, dimensionless

R_d - rate of total lime dissolution in a droplet, gmole/sec

R_p - rate of lime dissolution from a single lime particle, gmole/sec

Sc - Schmidt Number, dimensionless

Sh - Sherwood Number, dimensionless

t - time, sec

T - temperature, °C

t_{dust} - dust layer thickness, cm

U - gas velocity, cm/sec

v - relative velocity between gas phase and droplet, cm/sec

V - volumetric gas flow rate, cm^3/sec

w - moisture volume fraction in droplet, dimensionless

x - mole fraction, dimensionless

X - moisture content in solids, weight percent (dry basis)

X_c - critical moisture content, weight percent

X_e - equilibrium moisture content, weight percent

δ - film thickness, cm

ε - heat transfer correction factor, dimensionless

μ - viscosity, gm/cm-sec

ρ - density of gas phase, gm/cm^3

ρ_l - density of liquid phase, gm/cm^3

λ - latent heat of vaporization, cal/gm

η - fractional efficiency, dimensionless

Subscripts

d - droplet

g - gas phase

1 - liquid phase; lime

s - SO_2

SO_2 - SO_2

W - water

Superscripts

* - equilibrium

SECTION 11

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APPENDIX A

SPRAYMOD BASIC PROGRAM

```
10 REM  ************************************************************
20 REM
30 REM  Program  -  SPRAYMOD
40 REM
50 REM              Predicts Efficiency of Sulfur Dioxide
60 REM              Removal in a Spray Dryer using Sorbent Spray
70 REM
80 REM              Developed by Dr. Ashok S. Damle
90 REM                           Research Triangle Institute
100 REM
110 REM              For more information call (919) 541-5829
120 REM
130 REM  ************************************************************
140 CLS
150 PRINT"                        *** SPRAY DRYER MODEL ***"
160 PRINT"  ":PRINT"  "
170 PRINT"  Developed by : Dr. Ashok S. Damle"
180 PRINT"                 Research Triangle Institute"
190 PRINT"                 P.O.Box 12194"
200 PRINT"                 Research Triangle Park, NC 27709"
210 PRINT" " : PRINT" "
220 PRINT"  Menu Format for Data Input -"
230 PRINT"       A Default value is shown for each item in the"
240 PRINT"       bracket. Just hit enter if you don't want to change"
250 PRINT"       it. Otherwiae enter the new value." : PRINT" "
260 PRINT"                  Hit any key to start data input"
270 A$=INKEY$ : IF A$ = "" THEN 270
280 REM
290 REM  ************************************************************
300 REM
310 REM       START DATA INPUT
320 REM
330 REM  ************************************************************
340 REM
350 REM       Initialize Default Values
360 REM
370 TI=150:WI=10:AT=10:SI=800:DO=60:DP=4:TW=60:MF=1.0
380 MG=28.9:MS=74.0:SR=1.0:RR=0.0:OS=1:MR=0.3:SG=2.2
390 XC=50:XE=10:RK=0:RT=10:OP=2:NA$="":MT=0.1:PT=0.5:SC=0.01
400 REM
410 CLS
420 PRINT"":PRINT"":PRINT"":PRINT"":PRINT""
430 PRINT"   Do you want to read data from an "
440 PRINT"   Existing Datafile - ";:D$=""
450 INPUT D$:D$ = LEFT$ (D$,1)
460 IF D$ = "y" OR D$ = "Y" GOTO 470 ELSE 510
470 PRINT"":INPUT"   Enter Datafile Name - ";E$
480 OPEN "I", 1, E$
490 INPUT# 1, NA$,TI,WI,SI,MG,AT,SR,MF,RR,MR,RT,OP,DO,TW,OS,DP,MS,SG,XC,XE,RK,MT
,PT,SC
500 CLOSE 1
510 CLS
520 PRINT" ":PRINT" ":PRINT" ":PRINT" ":PRINT" ":PRINT" "
530 PRINT"       Dataset Name - ";NA$;"  ";:INPUT NA$
```

```
540 PRINT" ":PRINT" ":PRINT" ":PRINT" ":PRINT" ":PRINT" "
550 PRINT"                Do you want to redo this screen  ";
560 REDO$="":INPUT REDO$ : R$ = LEFT$ (REDO$,1)
570 IF R$ = "y" OR R$ = "Y" GOTO 510
580 CLS
590 PRINT" ":PRINT" "
600 PRINT"  Inlet Gas Conditions - ":PRINT" ":PRINT" "
610 PRINT"     Inlet Gas Temperature (";TI;"C) - ";:INPUT TI
620 PRINT" ":PRINT"     % Water Vapor in Inlet Gas (";WI;") - ";:INPUT WI
630 PRINT" ":PRINT"     Inlet SO2 Concentration (";SI;"ppm) - ";:INPUT SI
640 PRINT" ":PRINT"     Molecular Weight of Dry Inlet Gas (";MG;") - ";:INPUT MG
650 PRINT" ":PRINT" "
660 PRINT"                          Do you want to redo this screen ";
670 REDO$="":INPUT REDO$ : R$ = LEFT$ (REDO$,1)
680 IF R$ = "y" OR R$ = "Y" GOTO 580
690 CLS:PRINT"":PRINT"":PRINT"  Operaring Parameters -":PRINT" "
700 PRINT"     Approach to Saturation (";AT;"C) - ";:INPUT AT
710 PRINT"     Stoichiometric Molar Ratio (";SR;") - ";:INPUT SR
720 PRINT"     Mass Fraction of Reagent in Fresh Solids (";MF;") - ";:INPUT MF
730 PRINT"     Recycle Solids Ratio by Mass  (";RR;") - ";:INPUT RR
740 PRINT"     Mass Fraction of Reagent in Recycle (";MR;") - ";:INPUT MR
750 PRINT"     Residence Time of Gas Phase (";RT;"sec) - ";:INPUT RT
760 PRINT"     Gas Flow Pattern (Plug Flow=1,Backmixed=2) (";OP;") - ";:INPUT OP
770 PRINT" ":PRINT"                          Do you want to redo this screen ";
780 REDO$="":INPUT REDO$ : R$ = LEFT$ (REDO$,1)
790 IF R$ = "y" OR R$ = "Y" GOTO 690
800 CLS:PRINT"":PRINT"  Sorbent Properties - ":PRINT" "
810 PRINT"     Inlet Droplet Diameter (";DO;"um) - ";:INPUT DO
820 PRINT"     Inlet Droplet Temperature (";TW;"C) - ";:INPUT TW
830 PRINT"     Form of Sorbent (Slurry=1,Solution=2) (";OS;") - ";:INPUT OS
840 IF OS = 2 GOTO 860
850 PRINT"     Sorbent Particle Diameter (";DP;"um) - ";:INPUT DP
860 PRINT"     Molecular Weight of Sorbent (";MS;") - ";:INPUT MS
870 PRINT"     Density of Solid Sorbent (";SG;"gm/cm3) - ";:INPUT SG
880 PRINT"     Critical Moisture Content (";XC;"%) - ";:INPUT XC
890 PRINT"     Equilibrium Moisture Content (";XE;"%) - ";:INPUT XE
900 PRINT"     Dry Sorbent Reaction Rate Coeff. (";RK;"cm3/gmole/sec) - ";:INPUT
 RK
910 PRINT" ":PRINT"                          Do you want to redo this screen ";
920 REDO$="":INPUT REDO$ : R$ = LEFT$ (REDO$,1)
930 IF R$ = "y" OR R$ = "Y" GOTO 800
940 CLS:PRINT" ":PRINT" "
950 PRINT"  Program Parameters -":PRINT" ":PRINT" "
960 PRINT"     Maximum Allowable Time Step (";MT;"sec) - ";:INPUT MT
970 PRINT" ":PRINT"     Time between Output Printouts (";PT;"sec) - ";:INPUT PT
980 PRINT" ":PRINT"     Time Step Control Parameter (";SC;") - ";:INPUT SC
990 PRINT" ":PRINT" ":PRINT"                          Do you want to redo this screen ";
1000 REDO$="":INPUT REDO$ : R$ = LEFT$ (REDO$,1)
1010 IF R$ = "y" OR R$ = "Y" GOTO 940
1020 PRINT"":PRINT"                    Would you like to review complete data set";
1030 RE$="":INPUT RE$ : R$ = LEFT$ (RE$,1)
1040 IF R$ = "y" OR R$ = "Y" GOTO 510
1050 CLS:PRINT"":PRINT"":PRINT"":PRINT""
1060 PRINT"                    DATA INPUT DONE"
```

```
1070 PRINT"":PRINT"":PRINT"   Do you want to save this dataset - ";
1080 F$="":INPUT F$:F$=LEFT$ (F$,1)
1090 IF F$ = "y" OR F$ = "Y" GOTO 1100 ELSE 1230
1100 IF D$ = "y" OR D$ = "Y" GOTO 1110 ELSE 1190
1110 PRINT"":PRINT"   This dataset was read from - ";E$
1120 PRINT"":PRINT"   Do you want to replace this file with new data - ";
1130 G$="":INPUT G$:G$=LEFT$ (G$,1)
1140 IF G$ = "y" OR G$ = "Y" GOTO 1150 ELSE 1190
1150 KILL E$
1160 H$=E$
1170 GOTO 1200
1180 REM
1190 PRINT"":INPUT"   Enter a Datafile Name for This Dataset - ";H$
1200 OPEN "O", 1, H$
1210 PRINT# 1, CHR$(34);NA$;CHR$(34),TI,WI,SI,MG,AT,SR,MF,RR,MR,RT,OP,DO,TW,OS,D
P,MS,SG,XC,XE,RK,MT,PT,SC
1220 CLOSE 1
1230 CLS:PRINT"":PRINT"":PRINT""
1240 PRINT"         Check if Printer is On and Properly Connected"
1250 PRINT"":PRINT"              Hit Any Key to Continue"
1260 A$=INKEY$ : IF A$ = "" THEN 1260
1270 CLS
1280 REM
1290 REM        End Data Input
1300 REM
1310 REM  ************************************************************
1320 REM
1330 REM       Printout Data Values
1340 REM
1350 REM  ************************************************************
1360 REM
1370 LPRINT""
1380 LPRINT"                  *** SPRAY DRYER MODEL PREDICTIONS ***"
1390 LPRINT"":LPRINT"":LPRINT"   Input Dataset - ";NA$
1400 LPRINT"":LPRINT"   Inlet Gas Conditions - ":LPRINT""
1410 LPRINT"       Inlet Gas Temperature - ";TI;"C"
1420 LPRINT"       % Water in Inlet Gas - ";WI;"%"
1430 LPRINT"       Inlet SO2 Concentration - ";SI;"ppm"
1440 LPRINT"       Molecular Weight of Dry Inlet Gas - ";MG
1450 LPRINT"":LPRINT"   Operating Parameters - ":LPRINT""
1460 LPRINT"       Approach to Saturation - ";AT;"C"
1470 LPRINT"       Stoichiometric Molar Ratio - ";SR
1480 LPRINT"       Mass Fraction of Sorbent in Fresh Solids - ";MF
1490 LPRINT"       Recycle Solids Ratio by Mass - ";RR
1500 LPRINT"       Mass Fraction of Sorbent in Recycle - ";MR
1510 LPRINT"       Residence Time of Gas Phase - ";RT;"sec"
1520 IF OP=2 GOTO 1550
1530 LPRINT"       Flow System in Spray Dryer - Plug Flow"
1540 GOTO 1560
1550 LPRINT"       Flow System in Spray Dryer - Completely Backmixed"
1560 LPRINT"":LPRINT"   Sorbent Properties - ":LPRINT""
1570 LPRINT"       Inlet Droplet Diameter - ";DO;"um"
1580 LPRINT"       Inlet Droplet Temperature - ";TW;"C"
1590 IF OS=1 GOTO 1620
```

```
1600 LPRINT"      Form of Sorbent - Solution"
1610 GOTO 1640
1620 LPRINT"      Form of Sorbent - Slurry"
1630 LPRINT"      Sorbent Particle Diameter - ";DP;"um"
1640 LPRINT"      Molecular Weight of Sorbent - ";MS
1650 LPRINT"      Density of Solid Sorbent - ";SG;"gm/cm3"
1660 LPRINT"      Critical Moisture Content - ";XC;"%"
1670 LPRINT"      Equilibrium Moisture Content - ";XE;"%"
1680 LPRINT"      Dry Sorbent reaction Rate Coefficient - ";RK;"cm3/gmole/sec"
1690 LPRINT"":LPRINT"   Program Parameters - ":LPRINT""
1700 LPRINT"      Maximum Allowable Time Step - ";MT;"sec"
1710 LPRINT"      Time between Printouts - ";PT;"sec"
1720 LPRINT"      Time Step Control Parameter - ";SC
1730 LPRINT"":LPRINT""
1740 REM
1750 REM
1760 TI=TI+273.15:WI=WI*0.01:SI=SI*1E-6:LX=(MF+RR*MR)/(1+RR)
1770 DO=DO*1E-4:TW=TW+273.15:DP=DP*1E-4:XC=XC*0.01:XE=XE*0.01
1780 REM
1790 REM
1800 REM  ************************************************************
1810 REM
1820 REM         Calculate Adiabatic Saturation temperature
1830 REM
1840 REM  ************************************************************
1850 REM
1860 WP=WI*760
1870 HI=WP*18.0/MG/(760-WP)
1880 CS=0.24+0.45*HI
1890 T1=330.0
1900 L=597.64-0.275*(T1+TI-546.3)
1910 PS=EXP(20.426-5138.8/T1)
1920 YP=PS*18.0/MG/(760-PS)
1930 E1=TI-T1-(YP-HI)*L/CS
1940 IF ABS(E1)<0.1 GOTO 2110
1950 T2=T1+1
1960 L=597.64-0.275*(T2+TI-546.3)
1970 PS=EXP(20.426-5138.8/T2)
1980 YP=PS*18.0/MG/(760-PS)
1990 E2=TI-T2-(YP-HI)*L/CS
2000 IF ABS(E2)<0.1 GOTO 2120
2010 T3=T1-E1*(T1-T2)/(E1-E2)
2020 L=597.64-0.275*(T3+TI-546.3)
2030 PS=EXP(20.426-5138.8/T3)
2040 YP=PS*18.0/MG/(760-PS)
2050 E=TI-T3-(YP-HI)*L/CS
2060 IF ABS(E)<0.1 GOTO 2100
2070 IF ABS(E1)<ABS(E2) THEN T2=T1
2080 IF ABS(E1)<ABS(E2) THEN E2=E1
2090 E1=E:T1=T3:GOTO 2010
2100 TS=T3:GOTO 2130
2110 TS=T1:GOTO 2130
2120 TS=T2:GOTO 2130
2130 TE=AT+TS
```

```
2140 REM
2150 REM  ************************************************************
2160 REM
2170 REM          Overall Material and Energy Balances
2180 REM
2190 REM  ************************************************************
2200 REM
2210 PS=EXP(20.426-5138.8/TS)
2220 HS=PS*18.0/MG/(760-PS)
2230 TM=(TI+TE)/2
2240 L=597.64-0.55*(TM-273.15)
2250 CS=0.24+0.45*HS
2260 HO=HS-(TE-TS)*CS/L
2270 WO=HO/(HO+18.0/MG)
2280 PS=EXP(20.426-5138.8/TE)
2290 RH=WO*760/PS
2300 XU=RH*XE
2310 GW=1.01*(HO-HI)
2320 GS=SI*SR*MS*(1.0+RR)/MF/MG/(1.0-WI)
2330 ES=SR*(1.0+RR*MR/MF)
2340 XF=GW/(GW+GS)
2350 WO=0.52356*DO[3*SG/(1.0+(SG-1.0)*XF)
2360 TN=GW/XF/WO
2370 IF OS=2 GOTO 2420
2380 NP=((DO/DP)[3)*(1.0-XF)/(1.0+(SG-1.0)*XF)
2390 REM
2400 REM        Print Exit Gas Conditions
2410 REM
2420 LPRINT"   Exit Gas Temperature and Humidity ":LPRINT""
2430 LPRINT"      Exit Gas Temperature - ";TE-273.15;"C"
2440 LPRINT"      Adiabatic saturation Temperature - ";TS-273.15;"C"
2450 LPRINT"      % Water Vapor in Exit Gas - ";WO*100;"%"
2460 LPRINT"      % Relative Humidity of Exit Gas - ";RH*100;"%"
2470 LPRINT"":LPRINT""
2480 REM
2490 REM  ************************************************************
2500 REM
2510 REM      Initialize All Droplet and Gas Phase Variables
2520 REM
2530 REM  ************************************************************
2540 REM
2550 WW=XF*WO
2560 CM=WO*(1.0-XF)*LX/MS
2570 CI=WO*(1.0-XF)*(1.0-LX)/MS
2580 IF OP=2 GOTO 2660
2590 TG=TI:XW=WI:XS=SI:TD=TW
2600 GOTO 2870
2610 REM
2620 REM        For Backmixed Flow Option Assume SO2 removal
2630 REM        Efficiency of the Spray Dryer and Calculate
2640 REM        Outlet SO2 Concentration
2650 REM
2660 P1=0.0:P2=1.0
2670 IF ES<1.0 THEN P2=ES
```

```
2680 I=0
2690 REM
2700 REM        Start Trial and Error Sequence to Determine
2710 REM            SO2 Removal Efficiency
2720 REM
2730 I=I+1
2740 LPRINT"":LPRINT"    ***Trial ";I;" - Backmixed Model ***"
2750 P1=(P1+P2)/2.0
2760 IF I>1 GOTO 2820
2770 P1=P1*100
2780 PRINT" *ASSUMED EFFICIENCY OF THE SPRAY DRYER FOR THE"
2790 PRINT" *FIRST TRIAL IS ";P1;"% IF OK JUST HIT ENTER. IF NOT"
2800 INPUT" *ENTER NEW GUESS VALUE  - ";P1
2810 P1=P1/100
2820 LPRINT"   Assumed Spray Dryer Efficiency (%) = ";P1*100
2830 LPRINT"  "
2840 XS=SI*(1.0-P1):TG=TE:XW=WO:TD=TW
2850 REM
2860 REM
2870 T=0.0:TP=0.0:ST=0.0:D=DO:WD=WO:W=WW:X=XF:UT=0.0
2880 GOTO 4520
2890 REM
2900 REM  ************************************************************
2910 REM
2920 REM     Solve Material and Energy Balance Equations
2930 REM        for Droplet and Gas Phase
2940 REM
2950 REM  ************************************************************
2960 REM
2970 TM=(TD+TG)/2.0
2980 DS=2.951E-5*TM[1.5
2990 DW=5.062E-5*TM[1.5
3000 CT=1.2194E-2/TM
3010 CO=1.82E-7*TM+7.9E-6
3020 PW=EXP(20.426-5138.8/TD)
3030 XD=PW/760.0
3040 IF XD>1.0 THEN XD=0.9999
3050 IF X<XE THEN XD=XD*X/XE
3060 REM
3070 REM        Rate of Water Evaporation
3080 REM
3090 IF X > XU GOTO 3120
3100 WM=0.0 : FW=0.0
3110 GOTO 3150
3120 WM=6.2832*DW*CT*D*LOG((1.0-XW)/(1.0-XD))
3130 IF X<XC THEN WM=WM*(X-XU)/(XC-XU)
3140 REM
3150 RL=(CM-ABS(ST))/CM
3160 REM
3170 REM        Check for Dry or Wet state of droplet
3180 REM
3190 IF X<0.01 GOTO 3250
3200 IF X>XU*1.25 THEN GOTO 3350
3210 REM
```

```
3220 REM        Rate of SO2 Absorption in
3230 REM        Dry Particle Stage
3240 REM
3250 IF RL>0.001 GOTO 3270
3260 SM=0.0 : GOTO 3280
3270 SM=-RK*XS*CT*RL*CM
3280 FW=WM*18.0
3290 FS=SM*64.0
3300 GOTO 3770
3310 REM
3320 REM        Rate of SO2 Absorption in
3330 REM        Wet Particle Stage
3340 REM
3350 IF SM=0.0 GOTO 3390
3360 IF ABS(WM/SM)>3.0 GOTO 3390
3370 SM=-6.2832*DS*D*CT*XS
3380 GOTO 3400
3390 SM=XS*WM/(1.0-EXP(WM/DS/CT/D/6.2832))
3400 FW=WM*18.0
3410 FS=SM*64.0
3420 REM
3430 IF RL>0.001 GOTO 3470
3440 F=0.0:SM=0.0:FS=0.0
3450 GOTO 3770
3460 REM
3470 IF OS=2 GOTO 3770
3480 REM
3490 REM        Rate of Sorbent Dissolution    ( for OS=1 )
3500 REM
3510 VS=(1-X)/(1+(SG-1)*X)
3520 IF X<XC GOTO 3540
3530 LP=D*(VS/NP)[0.3333
3540 CL=1.88E-3-1.174E-5*(TD-273.15)
3550 F=-1.927E-7*TD*CL*LP*NP*64.0/MS
3560 REM
3570 REM        Calculate Film Thickness
3580 REM
3590 IF X<XC GOTO 3630
3600 BT=(0.973/VS[0.333)-1.0
3610 IF BT>1.0 THEN BT=1.0
3620 IF BT<0.5 THEN BT=0.5
3630 F=F/BT
3640 REM
3650 REM        Calculate Dissolution Rate Reduction due to
3660 REM          Product Precipitation and Inert Solids
3670 REM
3680 RF=1.0+(2.108*ABS(ST)+CI)/(CM-ABS(ST))
3690 IF RF>3.0 THEN RF=3.0
3700 F=F/RF
3710 REM
3720 IF F>FS THEN SM=F/64.0
3730 IF F>FS THEN FS=F
3740 REM
3750 REM        Rate of Heat Transfer
```

```
3760 REM
3770 IF WM=0.0 AND ABS(TG-TD)<0.2 THEN Q=0.0 :GOTO 3850
3780 EP=1.289*WM/CO/D
3790 IF EP<1.0E-5 THEN EP=1.0E-5
3800 Q=6.2832*CO*D*(TG-TD)*EP/(EXP(EP)-1)
3810 REM
3820 REM        Calculate Derivative Functions and
3830 REM           set Time Step DT
3840 REM
3850 F1=(Q-WM*10080)/WD
3860 F2=-FW-FS
3870 IF F1=0.0 THEN D1=1000:GOTO 3890
3880 D1=SC*(TD-273.0)/ABS(F1)
3890 IF F2=0.0 THEN D2=1000:GOTO 3910
3900 D2=SC*WD/ABS(F2)
3910 DT=D1
3920 IF D2<D1 THEN DT=D2
3930 IF DT>MT THEN DT=MT
3940 REM
3950 IF OP=2 GOTO 4110
3960 REM
3970 REM        Calculate Change in Gas Phase Variables
3980 REM           ( for OP=1 )
3990 REM
4000 CS=0.24+0.45*18.0*XW/MG/(1.0-XW)
4010 DG=Q*DT*TN/CS
4020 TG=TG-DG
4030 SD=SM*DT*TN*(1.0-XW)*MG
4040 XS=XS+SD
4050 DX=WM*DT*TN*(1.0-XW)*MG
4060 XW=XW+DX
4070 REM
4080 REM        Calculate Change in Droplet Variables during
4090 REM             Time Step DT
4100 REM
4110 T=T+DT
4120 ST=ST+SM*DT
4130 UT=-ST/CM
4140 TD=TD+F1*DT
4150 IF OS = 2 GOTO 4320
4160 REM
4170 REM       For Slurry Option (OS = 1)
4180 REM       Ca(OH)2 + SO2 + H2O = CaSO3.2H2O   Reaction is
4190 REM       Assumed and the Change in Weight of Droplet and
4200 REM       Water Content are Calculated Accordingly.
4210 REM
4220 WD=WD+F2*DT
4230 W=W-FW*DT+18.0*SM*DT
4240 IF W<0 THEN W=0
4250 GOTO 4360
4260 REM
4270 REM       For Solution Option (OS = 2)
4280 REM       Na2CO3 + SO2 = Na2SO3 + CO2   Reaction is Assumed
4290 REM       and the Change in Weight of Droplet and Water
```

```
4300 REM       Content are Calculated Accordingly.
4310 REM
4320 WD=WD+F2*DT+44.0*SM*DT
4330 W=W-FW*DT
4340 IF W<0 THEN W=0
4350 REM
4360 X=W/WD
4370 IF X<XC GOTO 4420
4380 D=((WD+(SG-1)*W)/SG/0.52356)[(1.0/3.0)
4390 REM
4400 REM        Check for Zero SO2 Removal During Last Time Step
4410 REM
4420 IF SM=0.0 GOTO 4520
4430 REM
4440 REM        Check for End Time and Printout Times
4450 REM
4460 IF T>RT GOTO 4520
4470 IF T<TP GOTO 2970
4480 TP=T+PT
4490 REM
4500 REM        Print Output Results
4510 REM
4520 LPRINT "TIME(SEC)=";T;"        GAS TEMP(C)=";TG-273.15
4530 LPRINT"DROP TEMP(C)=";TD-273.15;"        DROP DIAMETER(UM)=";D*1E4
4540 LPRINT"% WATER IN DROPLET=";X*100;"       WATER MOLE FRACTION IN GAS=";XW
OP=";UT*100
4560 LPRINT"   "
4570 IF T=0.0 GOTO 2970
4580 IF SM=0.0 GOTO 4640
4590 IF T>RT GOTO 4640
4600 GOTO 2970
4610 REM
4620 REM        Calculate SO2 Removal Efficiency
4630 REM
4640 EF=ES*UT
4650 IF OP=1 GOTO 4810
4660 P2=EF
4670 IF P2>1.0 THEN P2=1.0
4680 P3=ABS(P1-P2)
4690 IF P3<0.02 GOTO 4780
4700 IF I=1 THEN PA=P1:PB=(P1+P2)/2:GOTO 2730
4710 PC=(P1+P2)/2
4720 IF ABS(PC-PB)<ABS(PB-PA) GOTO 4760
4730 IF (PB-PC)/(PB-PA) > 0 THEN PC=(PA+PB)/2
4740 IF (PB-PC)/(PB-PA) < 0 THEN PC=1.5*PB-PA/2
4750 P2=2.0*PC-P1
4760 PA=PB:PB=PC
4770 GOTO 2730
4780 EF=(P1+P2)/2:XS=SI*(1.0-EF)
4790 REM
4800 REM
4810 LPRINT"% SO2 REMOVAL EFFICIENCY=";EF*100
4820 LPRINT"EXIT SO2 CONCENTRATION(PPM)=";XS*1E6
```

```
4830 LPRINT" "
4840 LPRINT" "
4850 LPRINT" "
4860 REM
4870 REM        Check for Remaining Datasets
4880 REM
4890 INPUT"NO. OF CASES REMAINING?  ";N
4900 IF N>0 GOTO 140
4910 STOP
```